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# Sludge Dewatering Design Manual

## Research Report No. 72

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**Research Program for the Abatement of Municipal Pollution under Provisions of the Canada-Ontario Agreement on Great Lakes Water Quality**

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Sludge dewatering design manual / Campbell, H. W.

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# SLUDGE DEWATERING DESIGN MANUAL

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by

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FISHERIES AND ENVIRONMENT CANADA

RESEARCH PROGRAM FOR THE ABATEMENT
OF MUNICIPAL POLLUTION UNDER THE
PROVISIONS OF THE CANADA-ONTARIO
AGREEMENT ON GREAT LAKES WATER QUALITY

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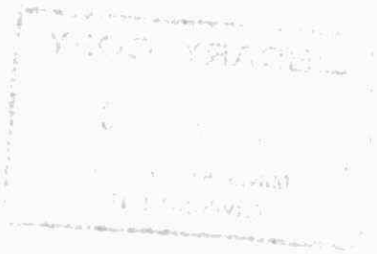

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## ABSTRACT

The dewatering manual integrates bench and pilot scale data generated at the Wastewater Technology Centre with information assembled from the literature, with a view to presenting a practical step-by-step procedure for estimating the size and cost of sludge dewatering installations.

Sludge characterization tests are identified, along with a discussion of each test and its potential use for design purposes.

A procedure for estimating the quantities of primary, waste activated and digested sludge, including the contribution of sludge from phosphorus removal processes, is presented.

Process design procedures for gravity thickening, dissolved air flotation, centrifugation, vacuum filtration and pressure filtration are presented. The procedure for each dewatering method includes a description of the unit process, a discussion of applicable bench or pilot plant tests, a suggested method of scale-up and an example problem.

Cost data (December, 1975) for the major pieces of equipment required for each process have been obtained from equipment suppliers and are reported as a function of an appropriate design parameter. Operating and maintenance costs collected from the literature are summarized. A methodology for developing cost estimates is presented and illustrated with example calculations.

## RÉSUMÉ

Le manuel de déshydration des boues rassemble des données produites par le Centre de la technologie des eaux usées ainsi que d'autres, trouvées dans les publications spécialisées. On y décrit en outre, par étapes, une méthodes permettant d'évaluer l'importance et le coût des installations de déshydratation.

Le manuel comprend aussi:

- des tests de caractérisation des boues, tous commentés et dont on expose l'utilisation éventuelle comme critères de conception;
- une méthode permettant d'évaluer le volume des boues du traitement primaire, des boues activées et des boues digérées, ainsi que l'apport en boues de la déphosphatation;
- des méthodes de conception des procédés d'épaississement par décantation, de flottation par infection d'air, de centrifugation, de filtration sous vide et sous pression avec, dans chaque cas, la description du procédé, des commentaires sur l'application des tests de laboratoire ou des installations pilotes, une méthode recommandée d'extrapolation de même qu'un exemple type;
- des données sur le coût des principales pièces d'équipement, obtenues des fournisseurs et présentées en fonction d'un paramètre approprié de conception;
- un résumé des frais d'entretien et d'exploitation compilés à partir des publications spécialisées, ainsi qu'une méthode d'estimation des coûts accompagnée d'exemples de calculs.

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# 1 INTRODUCTION

The design of sludge treatment facilities has traditionally received much lower priority than the design of treatment systems for the liquid portion of the waste. This may have been due partially to the fact that until the liquid treatment facility was in operation there was no sludge, and to the fact that while liquid discharges were regulated by government standards, sludge disposal was not. It is still a common occurrence to find flow diagrams of waste treatment systems which end with an arrow and the caption "to sludge disposal", even though sludge handling and disposal costs may represent up to 50% of the total cost of waste treatment. In view of this, it is essential that sludge treatment design procedures generate data which will allow meaningful comparisons between various alternatives, and lead to the selection of the optimum system.

The potential impact of nutrient removal systems on the sludge handling requirements of conventional sewage treatment plants was recognized by the Canada-Ontario Agreement on Great Lakes Water Quality as one of the areas requiring further research. As a result a "Sludge Treatment Process Development Program" was initiated to characterize sludges, and to examine the control variables and scale-up techniques for several conventional dewatering installations. The work carried out under this project (Campbell et al, 1975) was instrumental in defining three areas of sludge dewatering where the information as presented in the literature was inadequate.

Most of the techniques and tests used to characterize sludges are reported in the literature, along with typical results for various types of sludges. In contrast the interpretation and practical application of these tests as aids in the selection and design of a dewatering process is not well documented.

The design of dewatering installations should be based on data developed from either bench tests or pilot plant operation. The methodology whereby a designer can perform bench or pilot plant tests, interpret the data, and systematically develop specifications for dewatering equipment must be assembled from several literature sources for each process.

Since sludge dewatering equipment represents a major portion of the total cost of a sewage treatment plant a preliminary estimation of this

cost is essential. Developing a cost estimate on the basis of information obtained from the literature is a time consuming and frustrating task. Discrepancies between literature sources as to what components are included in a unit cost, such as capital investment, can easily result in cost estimates for the same system differing by a factor of 3 to 4.

The preparation of a sludge dewatering design manual, by integrating data generated at the Wastewater Technology Centre with information assembled from the literature, was undertaken to satisfy the following objectives:

1. *to identify meaningful sludge characteristics and discuss their significance and practical application to the design of various dewatering processes;*

2. *to outline a systematic process design methodology proceeding from bench tests, to pilot plant operation, to full scale design;*

3. *to develop a cost data base for sludge dewatering equipment which has common items included in each unit cost at a common time base.*

The information presented in this manual is not intended for detailed plant design or cost analysis. The aim of this manual is to provide a methodology by which it is possible to estimate full scale dewatering requirements on the basis of experimental data, and develop an initial cost estimate for either comparing alternative systems or evaluating the economic feasibility of any specific system.

## 2 SLUDGE CHARACTERIZATION

The measurement of sludge characteristics serves a three-fold purpose. It provides a basis for comparing sludges from different sources within a waste treatment plant, from the same source in different plants, and from different types of waste treatment systems. In some cases it may be used to identify at a very early stage whether or not a particular dewatering process is applicable. Frequent monitoring of a sludge characteristic within a plant may pinpoint significant changes which will affect the dewaterability of a sludge.

The utility of any sludge measurement depends upon a knowledge of the variability of a sludge. Sludge sampling methodologies are available for determining the number of samples required to obtain an estimate of the mean of a variable within a specified level of confidence (Monteith and Stephenson, 1978). The degree of variability exhibited by a sludge dictates to a large extent the magnitude of safety factor which must be incorporated into the final design of sludge handling and dewatering equipment.

### 2.1 Solids Analysis

The test most frequently used in sludge analysis is the determination of the total solids concentration. The addition of conditioning chemicals is calculated on the basis of weight of chemical per unit weight of dry solids in the sludge, but the metering of the chemical solution is on a volumetric basis. In order to maintain an accurate dosage it is essential to know what the average solids concentration in the sludge is, and the degree of variation which can be expected. Typical concentrations of sludge may range from 0.5% to 2.0% w/w for waste activated sludge, and 4% to 6% w/w for primary and digested sludges. Some chemical sludges, especially those with high lime content, may be even higher. The solids concentration of the sludge is an important factor in determining possible methods of dewatering. Vacuum filtration generally requires a minimum of 4% solids in the feed, whereas dissolved air flotation is seldom used to concentrate sludges with initial solids levels of greater than 2% or 3%.

The density of the solids fraction is of particular interest to those processes which depend upon a solid-liquid density difference to

accomplish separation (gravity thickening, dissolved air flotation, and centrifugation). Reported values of dry solids densities for municipal sludges ranged from 1.4 to 2.1 g/cm$^3$, while the bulk density of the suspension ranged from 1.0032 to 1.054 g/cm$^3$ (Campbell et al, 1975). The measured densities were more dependent upon the chemical added for phosphorus removal than on the type of sludge (primary, secondary, or digested). Lime addition consistently resulted in higher densities whereas the effect of adding alum or ferric chloride did not follow a definite trend. Since the floc density is usually very close to the bulk density of the solution, conditioning is frequently used to increase the effective density of the floc aggregates and thus increase sedimentation rates.

## 2.2 Particle Size

The particle size distribution of sludge particles has a significant effect on dewatering performance but it is a characteristic which is very difficult to measure accurately. A problem arises in trying to define what constitutes a particle. In a well flocculated sludge, smaller particles are loosely bound together to form an aggregate floc. Whether or not the aggregate has sufficient structural strength to stay together as a unit throughout the dewatering process should be the determining factor in deciding what the functional size unit is.

Several methods have been used with varying success to obtain particle size distributions (Vesilind, 1974c). Wet screening of the sludge produces a size distribution on the basis of mass. Particle size distribution can be determined using a microscope, but the procedure is time consuming and requires interpretation as to what constitutes a particle.

The use of a Coulter Counter for particle size measurement requires dilution of the sludge with an electrolyte. The mixture is stirred in a vessel, a sample drawn through an aperture, (30 to 2000 μ diameter) and the pulses created by the sludge particles measured. By assuming that the particles are spheres it is possible to obtain a distribution on the basis of volume frequency. A typical particle size distribution diagram is shown in Chapter 3 (Figure 3-1). The breakup of the floc due to a high shear zone around the aperture may unrealistically weight the distribution towards

smaller particles, but the recording of particle diameters as high as 1000 μ indicates that some of the larger flocs have sufficient strength to withstand the shear forces. Although the data available on this type of measurement is very limited, the majority of sludge particles appear to be less than 100 μ in diameter. The most important aspect of the distribution is the percentage of particles which are in the colloidal range. It is these particles which have the lowest settling velocity, tend to blind filter cloths and exhibit the greatest chemical demand. Any procedure which can agglomerate colloidal particles into larger flocs will improve dewaterability. The mechanisms by which conditioning methods accomplish this, will be discussed in a later section.

At the present time, use of a Coulter Counter is considered to be primarily a research tool because of its high capital cost and the high degree of operator skill required.

## 2.3 Rheology

The rheological or flow properties of sludges are important to the design of pumping, mixing and dewatering equipment. The viscosity of a sludge, defined as the rate of displacement of a fluid upon application of a given shear force, determines in part the energy required to move the slurry through a pipe. Most fluids are classified as Newtonian fluids in that the shear stress is proportional to the displacement. In sludge the suspended particles interfere with the free flow of the dispersion medium to a degree that is dependent on the rate of shear. These slurries are non-Newtonian fluids and generally exhibit either plastic or pseudo-plastic flow behaviour. They possess an initial characteristic yield stress after which shear stress is proportional to shear rate, in the case of plastic materials. For pseudo-plastic materials the slope of the shear stress versus shear rate curve (viscosity) decreases with increasing shear rate.

The rheology of sewage sludges is further complicated by the fact that most sludges are also thixotropic, meaning the flow properties are time dependent. A rheogram (Figure 2-1) of a typical sludge, showing the curves produced by a rotating viscometer during the increasing and decreasing rate of shear cycles, indicates that the rheology of the sludge has been altered during

the initial phase of the test. The displacement of the two curves is referred to as a hysteresis loop and is a measure of the degree of thixotropy exhibited by a sludge.

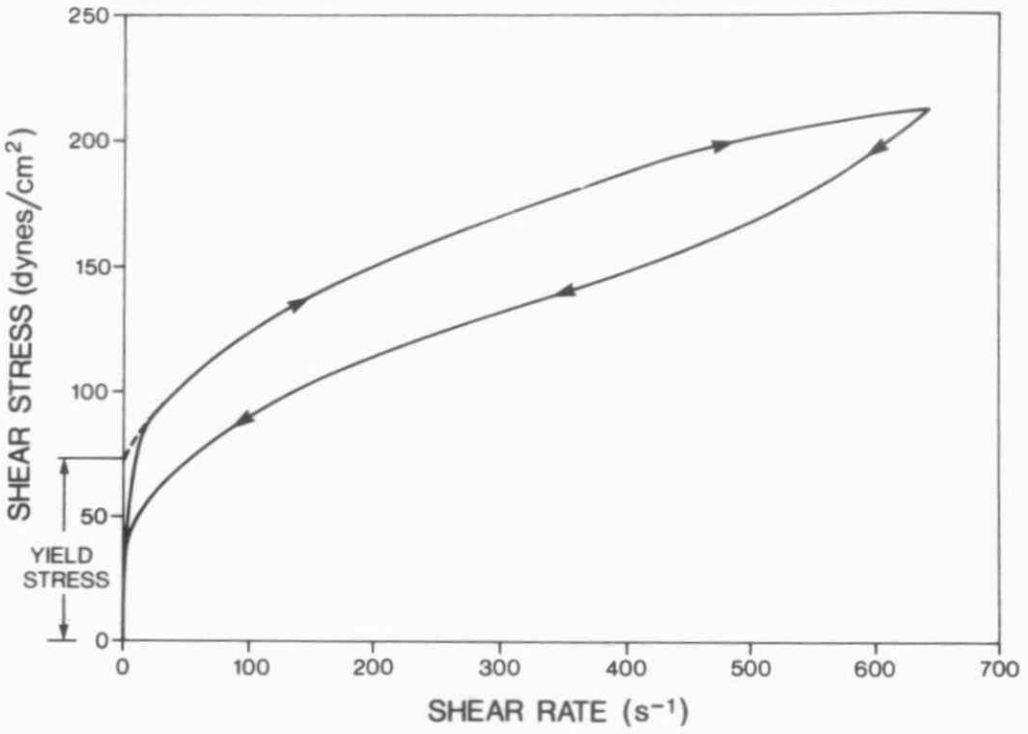


FIGURE 2-1. TYPICAL RHEOGRAM ILLUSTRATING THIXOTROPY AND YIELD STRESS

The rheology of sludges is greatly influenced by the suspended solids concentration. As the concentrations increase, both yield stress and viscosity increase for a particular sludge (Figure 2-2). When comparing the rheology of different sludges there are other factors which are more important than solids concentration. In Figure 2-3 typical rheograms are presented for a waste activated, a lime treated digested and a ferric chloride treated digested sludge. Although the solids concentration of the lime sludge is higher than that of the ferric chloride sludge (9.7% compared to 6.6%) the

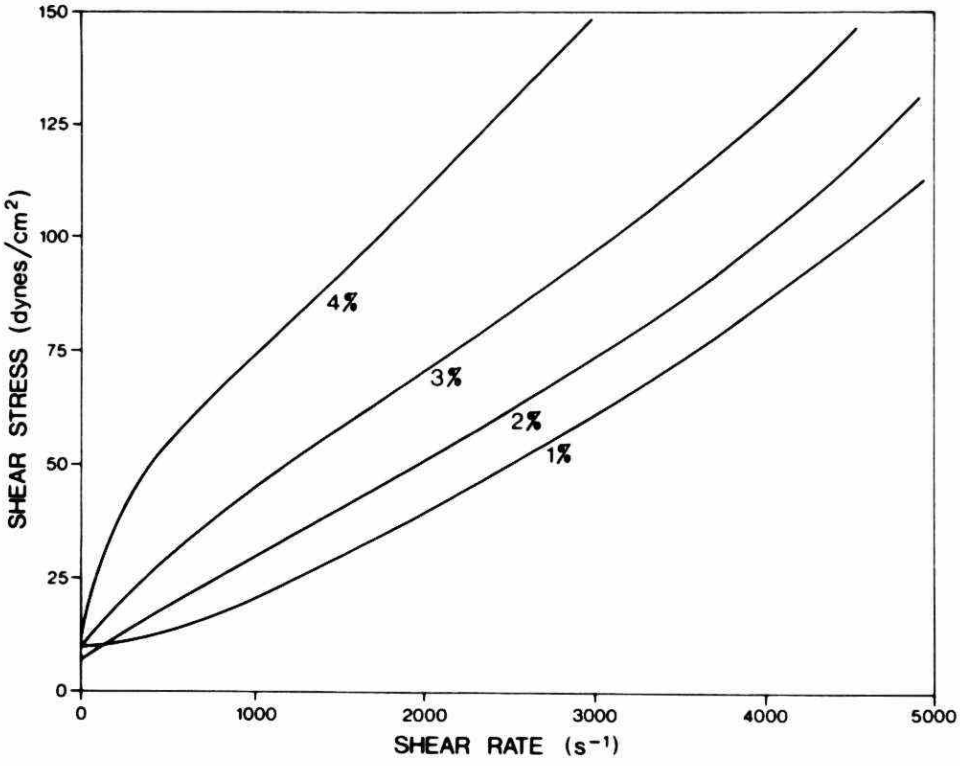


FIGURE 2-2. EFFECT OF SOLIDS CONCENTRATION ON RHEOLOGY OF WASTE ACTIVATED SLUDGE

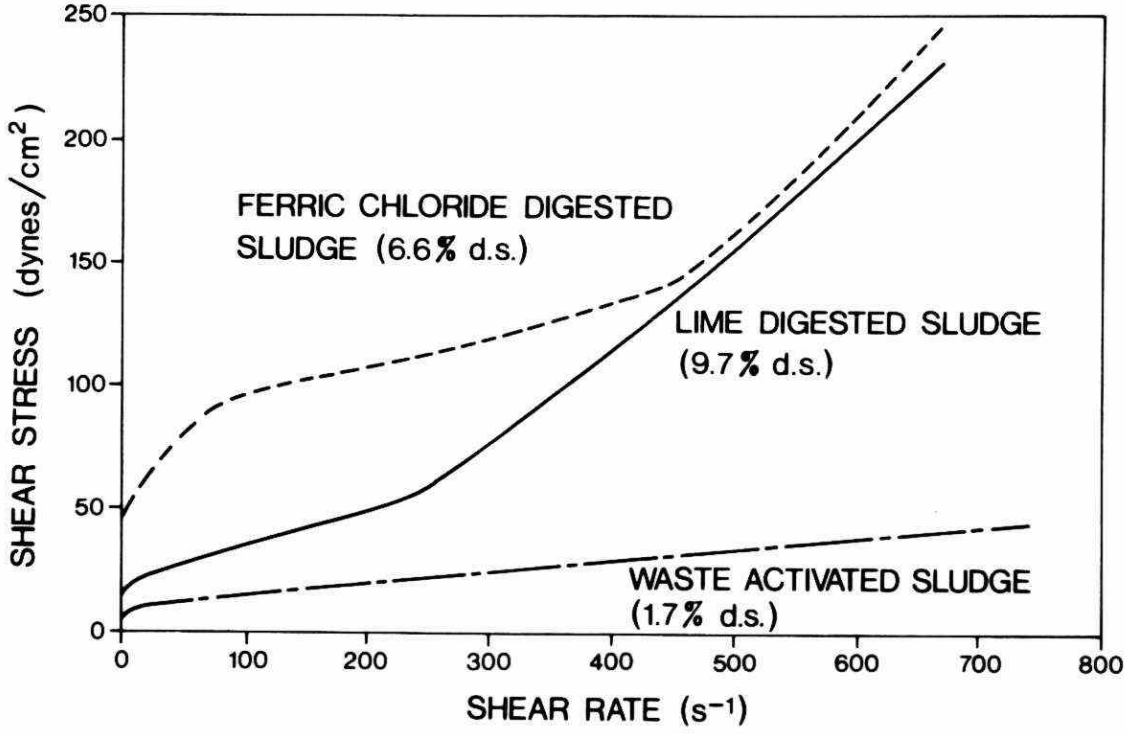


FIGURE 2-3. COMPARISON OF RHEOLOGICAL PROPERTIES OF THREE SLUDGES

yield stress of the lime sludge is lower. In this case the rheological differences are governed by particle characteristics as opposed to particle concentration.

Rheological measurements cannot at present be used to generate design data for sludge dewatering processes. The measured parameters (viscosity and yield stress) will indicate unusual sludges which may require special treatment. The greatest potential use of rheograms appears to be in the area of sludge conditioning. The relationship between chemical conditioning and rheology will be discussed in Chapter 3.

## 2.4 Zeta Potential

One method of promoting sludge flocculation is to reduce the electrical charge on a particle by chemical addition to move the zeta potential towards the isoelectric point (Riddick, 1968). Sewage sludges generally exhibit a net negative charge in the range of -10 to -20 millivolts which does not appear to depend upon the type of sludge (primary, secondary or digested). One application of the zeta potential is as a means of determining the optimum chemical dosage required. Theoretically this can be accomplished by monitoring the zeta potential with increasing chemical dosage and assuming the optimum conditioning level will occur when the zeta potential equals zero. This approach is not entirely satisfactory for use with sewage sludges. As conditioning levels increase, particle size increases, and the sedimentation rate of the particles increases to a point where observation of the particle and subsequent zeta potential determinations become extremely difficult.

The relatively narrow range of zeta potential exhibited by sewage sludges is not consistent with the wide range of chemical conditioning levels required, if the reduction of zeta potential is a governing factor. It is not uncommon for sludges with similar zeta potentials to require chemical addition levels differing by a factor of two, in order to achieve similar degrees of filterability. Sludges (primarily industrial) which exhibit very high zeta potentials in the range of -50 to -80 millivolts can be expected to be extremely difficult to dewater. In these cases electrical charge reduction would be a primary objective of any sludge conditioning method. The use of

zeta potential as a means of optimizing chemical addition has been used with some success in dewatering dilute waste activated sludge in laboratory studies (Roberts and Olsson, 1975) but practical application of the technique has not been reported.

## 2.5 pH and Alkalinity

The pH of most municipal sludges, with the possible exception of those which have had chemicals added for phosphorus removal, is near neutral. The pH of chemical-biological sludges is more dependent upon the sources within the treatment plant (primary, secondary, digested) than on the chemical added. Lime addition to a primary clarifier can result in a raw primary sludge with a pH of 9 to 11, but if this sludge is subsequently digested the pH normally drops to about 7. The primary concern for dewatering is that the pH of the sludge is in the optimum range for whichever conditioning chemical is being used. The performance of polymers is relatively insensitive to the pH range normally found with sewage sludges, but may deteriorate at high levels ($\simeq$11) resulting from lime addition. Sludge conditioning for vacuum filtration is often controlled by reducing the pH to $\simeq$4 with ferric chloride and then raising it to $\simeq$11 by the addition of lime.

The bicarbonate alkalinity varies with the type of sludge. Primary and secondary sludges may have relatively low alkalinities of a few hundred mg/ℓ while digested sludges may be as high as 4000 to 5000 mg/ℓ. The effect of alkalinity on polymer conditioning is not documented but high alkalinities represent an increased chemical demand when ferric chloride is used as the conditioner. Elutriation is often practiced to reduce the alkalinity of digested sludge prior to chemical conditioning (Vesilind, 1974c).

## 2.6 Sludge Volume Index

Sludges may be characterized by their settling characteristics. One measure of these characteristics is the Sludge Volume Index (SVI). This can be determined by allowing sludge to settle for 30 minutes in a one litre graduated cylinder. The SVI is calculated by the following equation:

$$SVI = \frac{\text{volume of settled sludge (m}\ell)}{\text{initial suspended solids concentration (g/}\ell)} \qquad (2\text{-}1)$$

This test is most applicable to gravity thickening of waste activated sludge where a SVI of $\leq$100 denotes a well settling sludge. Due to the inherent problems of conducting a settling test in a small cylinder (1 litre), the results should only be used in a qualitative manner. Sludges which exhibit a high SVI seldom concentrate well in a gravity thickener and some other process, such as dissolved air flotation, may be preferable. SVI determination is a quick and effective method of indicating gross daily changes in the settling characteristics of a specific sludge. It is most useful as an operators tool for day to day control of a treatment process.

## 2.7 Specific Resistance

Specific resistance is the most widely reported parameter describing the relative dewaterability of sludges. It is defined as the pressure difference required to produce a unit rate of filtrate flow of unit viscosity through a unit weight of cake. Values for specific resistance are determined using a Buchner funnel (Eckenfelder and Ford, 1970). A sample of sludge is filtered at a specific pressure and the volume of filtrate collected is monitored with respect to time. The ratio of time/volume is plotted as a function of volume of filtrate and should result in a straight line. The specific resistance is calculated from the following equation:

$$r = \frac{2 P A^2}{\mu C} b \qquad (2\text{-}2)$$

where: $r$ = apparent specific resistance (m/kg).
$P$ = filtration pressure ($N/m^2$).
$A$ = filtration area ($m^2$).
$\mu$ = filtrate viscosity ($N\text{-}s/m^2$).
$C$ = mass of dry suspended solids per unit volume of liquid in the sludge ($kg/m^3$).
$b$ = slope of time/volume versus volume graph, determined from the Buchner funnel test ($s/m^6$).

The value of r determined from the above equation is the apparent specific resistance and is the value usually reported in the literature.

The value of the "true" specific resistance r* can be calculated from the same equation with the exception that C* is used instead of C. C* is defined as the mass of dry suspended solids deposited on the filter cake per unit volume of filtrate obtained. Due to the difficulty of determining values for C* experimentally, it is generally approximated by C which is given by the following equation:

$$C = \frac{w_s}{(1-w_s) / \rho_L} \qquad (2\text{-}3)$$

where: $w_s$ = percentage by weight of dry suspended solids in the sludge (expressed as a fraction e.g., at 4%, $w_s = 0.04$).

$\rho_L$ = density of liquid ($kg/m^3$).

(For sewage sludges $\rho_L$ is normally assumed to be the same as that for water.)

It should be noted that many values for specific resistance quoted in the literature in terms of $s^2/g$ are theoretically incorrect (Gale, 1967; Tebbutt, 1970; Rushton and Rushton, 1973). This arises from incorrectly expressing pressure in terms of mass per unit area ($g/cm^2$) as opposed to force per unit area ($dyne/cm^2$). The proper units for specific resistance are m/kg and this convention is adhered to in most recent publications. Values reported in $s^2/g$ can be converted to m/kg by multiplying $s^2/g$ by $9.81 \times 10^3$.

The specific resistance provides a quantitative measure of the filterability of a sludge. A value of $0.1 \times 10^{13}$ m/kg (at 49 $kN/m^2$) or less is normally required for filtration to be an efficient method of dewatering (Gale, 1971b). Some typical values for the specific resistance of wastewater sludges which have been treated for phosphorus removal are shown in Table 2-1. All of these sludges would require conditioning to lower the specific resistance before an acceptable yield could be obtained on a vacuum filter. The specific resistance also gives a qualitative indication of how well a sludge will dewater using other processes, such as centrifugation (the lower the specific resistance, the easier it will centrifuge).

The measurement of specific resistance is widely used as a means of monitoring the effects of chemical conditioning, both in terms of chemical selection and optimum dosage. It may also be used to give a theoretical prediction of the filter yield.

TABLE 2-1. TYPICAL VALUES OF SPECIFIC RESISTANCE (x $10^{13}$ m/kg)[1]

| Type of Sludge \ Type of Chemical Addition | Lime | Ferric Chloride | Alum | No Chemical Addition |
|---|---|---|---|---|
| Raw Primary | 4.5 | 25.0 | 29.0 | -- |
| Waste Activated | -- | 1.9 | 0.4 | 1.5 |
| Mixed Anaerobic Digested | 4.5 | 35.3 | 26.4 | -- |

[1]Measured at 91.2 kN/$m^2$.
(From Campbell et al, 1975).

## 2.8 Compressibility

The coefficient of compressibility can be determined from the Buchner funnel test. Several tests at different pressures are conducted and the specific resistance plotted against the pressure on log-log paper. The slope (S) of the line obtained is the coefficient of compressibility. For incompressible materials such as sand S = 0 and the value increases as the material becomes more compressible. Values reported for sewage sludges range from 0.5 to 1.1 (Eckenfelder, 1970). The compressibility of a sludge increases as the degree of flocculation increases (Coakley and Wilson, 1971). The coefficient of compressibility may be used to calculate the specific resistance at pressures other than those used for the Buchner funnel test according to the following equation (Vesilind, 1974c):

$$r_2 = r_1 \left( \frac{P_2}{P_1} \right)^S \qquad (2\text{-}4)$$

where: S = coefficient of compressibility.
$r_1$ = specific resistance at pressure $P_1$.
$r_2$ = specific resistance at pressure $P_2$.

A knowledge of the compressibility of a sludge is useful in determining the pressure to be used in a process such as pressure filtration. High pressures are normally avoided when pressing a very compressible sludge, since the sludge tends to compress or compact to the point where liquid flow through the cake is no longer possible.

## 2.9 Capillary Suction Time

The capillary suction time test (CST) was developed as an alternative to the Buchner funnel test for measuring the filterability of sludge. The Buchner funnel test requires several pieces of equipment, and can be extremely time consuming when dealing with unconditioned sewage sludges. The additional time required to analyze the data and calculate specific resistance makes the use of this test impractical for process monitoring. The CST can be determined quickly and easily in the field, produces results which can be interpreted directly, and provides a basis for immediately evaluating changes in the sludge processing system.

The CST is determined by pouring a sample of sludge into a reservoir resting on an absorbent filter paper, where the filtrate is drawn out of the reservoir by the capillary action of the paper. The liquid saturates the paper and spreads outwards in a circular pattern at a rate determined by the properties of the paper, the viscosity of the filtrate and the filterability of the sludge. The time taken for the liquid front to move a specified distance is measured with electrical probes and a timer. The measured time in seconds is the capillary suction time. Complete details concerning equipment and procedures used are available in the literature (Baskerville and Gale, 1968).

Capillary suction time can be correlated with specific resistance by conducting a series of Buchner funnel tests and CST's on the same test batches of sludge. It has been found that the product of specific resistance and solids concentration gives a much better correlation with CST and that it is advisable to determine the correlation for each specific sludge (Baskerville and Gale, 1968). Figure 2-4 shows the correlations obtained for two digested sludges and one primary sludge, and also gives an indication of the wide range of CST values which are found with sewage sludges.

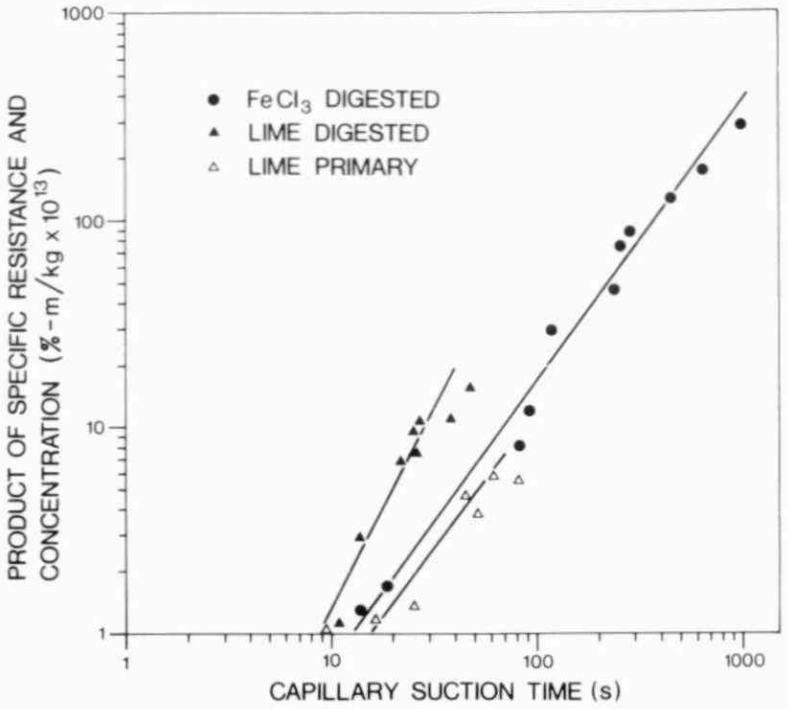


FIGURE 2-4. CORRELATION OF CST AND SPECIFIC RESISTANCE FOR SLUDGES FROM PHOSPHORUS REMOVAL FACILITIES

Due to the minimal time required for each test, the CST has been used extensively in the selection of sludge conditioning agents, optimum addition levels of these materials and the effect of variables such as mixing time and speed on sludge filterability (Baskerville and Gale, 1968). The simplicity of the test also makes it ideal for process control applications. Chemical addition for sludge conditioning may be controlled on the basis of achieving a predetermined CST value at some point within the process.

# 3 SLUDGE CONDITIONING

Sewage sludges possess several characteristics which directly inhibit the rate and performance (degree of clarification and thickening) achieved by any solid-liquid separation process. The presence of colloidal particles increases the specific resistance of the sludge and adversely affects sedimentation processes. Most sewage sludges exhibit a net negative charge. This charge makes the particles tend to repulse each other and thus resist agglomeration into larger particles. A large percentage of the sludge solids have a film of water, normally referred to as bound water, firmly attached to the particles. The retention of this water can result in low cake solids after solid-liquid separation. The objective of sludge conditioning is to alter by physical or chemical means any of the properties outlined above so as to increase the ease with which water can be removed from the sludge.

## 3.1 Physical Methods

Three common methods of physically conditioning sludges are freezing, heat treatment and the addition of inorganic admixtures.

Slow freezing of sludge slurries can result in a sludge which dewaters extremely well by gravity on subsequent thawing. A typical study showed that a conventional waste activated sludge could be concentrated from 0.6% to 16% using a freeze-thaw cycle, while digested sludge could be thickened from 5.2% to 25% (LeClair and Stickney, 1975). The disadvantages of the system are related to the quality of the effluent and the cost of the process. The effluent has low suspended solids but a high $BOD_5$ which would exert a considerable biological load if recycled back to the treatment plant. Due to the energy requirements it is doubtful if the process could be economically justified unless freezing could be accomplished by natural means.

Subjecting sludge to high levels of heat and pressure breaks down the gel-like structure of most sludges and allows water to escape. The resulting slurry can be readily dewatered without chemical conditioning (Vesilind, 1974c). The effluent from the dewatering process contains the contents of the ruptured cells and subsequently contributes significantly to

the organic loading on the treatment plant. The combination of high temperature and pressure requires specialized equipment which is normally supplied as a manufacturer's package. Due to the safety hazards associated with this type of equipment, operation and maintenance is generally more critical than would be expected with other conditioning systems.

The use of inorganic admixtures involves the addition of materials such as fly ash, incinerator ash, or diatomaceous earth to the sludge in sufficient quantities to produce a porous lattice. The resulting mixture has improved filtering characteristics and is usually less compressible than the sludge alone. Admixtures have been used extensively to condition sludge for pressure filtration and to a lesser extent for vacuum filtration. The decrease in compressibility is particularly advantageous for these two processes. Since the ratios of admixture to dry sludge solids typically range from 0.5:1 to 2:1 the weight and volume of the admixture must be considered for any ultimate disposal system. For example, if the sludge is to be trucked to a landfill site, a considerable increase in cake solids concentration is required, to compensate for the amount of admixture added. At a given cake solids concentration the addition of inorganic materials reduces the overall calorific value of the sludge for incineration (Schroeder, 1977). Care must also be taken to ensure that the weight of admixture has been accounted for when assessing the performance of a conditioning system on the basis of dry solids yield.

## 3.2 Chemical Methods

Chemical conditioning can be accomplished by using either organic or inorganic chemicals. Inorganic chemicals are usually either metal salts or lime or a combination of both. The most commonly used metal salt is ferric chloride, but ferrous sulphate, aluminum chlorohydrate and alum have also been used. In an aqueous solution metal salts hydrate to form free water-metal ion complexes and metal hydroxide precipitates. The coagulation-flocculation process occurs by the following three methods:

1. Neutralization of charged particles by the water-metal ion complexes.

2. Adsorption of the hydroxide on the particles and subsequent bridging between particles.
3. Enmeshment of the particle in precipitating metal hydroxide. The particle may act as a nucleus for this precipitation.

The mechanisms described are pH dependent since some metal hydroxides are soluble at both low and high pH's. If a sludge is highly buffered, such as digested sludge which normally has a high alkalinity (2000 to 6000 mg/ℓ as $CaCO_3$), the required coagulant dose of metal salt is very high. A common method of reducing buffering capacity is to add lime which precipitates the carbonate as calcium carbonate. This results in the proper pH range for optimum use of the metal salt and some of the sludge particles are enmeshed in the precipitated calcium carbonate. This situation is highly desirable because calcium carbonate has much better filtering characteristics than the metal hydroxides.

Most of the organic materials used in sludge conditioning are high molecular weight, water soluble organic polymers. Depending upon the type of ionizable group (e.g., carboxyl, amino, sulphonic) the polymers are termed cationic (positive charged groups), anionic (negative charged groups) or ampholytic (both positive and negative groups). If the polymer does not ionize in water it is referred to as non-ionic. The mechanisms by which polymers condition sludge are similar to those described for the metal salts, and consist essentially of charge neutralization, surface adsorption and inter-particle bridging. Due to the net negative charge exhibited by most sewage sludges, cationic polymers have enjoyed the most widespread use in the wastewater field. A vast array of commercial polymers which differ in the type of ionizable group, the molecular weight, the degree of hydrolysis, the physical characteristics (powder or liquid) and the cost, are presently available.

## 3.3 Evaluation of Chemical Conditioners

The ultimate test of the efficiency of any chemical conditioner must be based on the actual performance of full scale dewatering equipment. Due to the large number of products available it is essential that a simple,

quick method of preliminary screening of chemical conditioners be employed. The choice between organic and inorganic conditioners can sometimes be made on the basis of the available dewatering process. Due to its abrasive qualities lime is seldom used as a conditioner for centrifugation, which is highly susceptible to abrasion. Polymers have been used almost exclusively for this type of dewatering process. A combination of lime and ferric chloride has traditionally been used to condition sludges for vacuum filtration. Polymers are receiving increasing attention for application to vacuum filtration but the selection of a polymer is very specific for a given sludge. Dissolved air flotation normally employs either alum or polymer since the high density of a material such as lime would greatly hinder flotation. Lime, incinerator ash or fly ash have been the most common materials used to condition a sludge for pressure filtration. Although the choice between polymers and inorganic conditioners for any process is usually based on performance and cost, it should be noted that the addition of a polymer will normally result in smaller additional volumes of dewatered sludge.

The need for a test which quickly screens conditioners is most evident when attempting to choose a suitable polymer. Due to the selective nature of many polymers it may be necessary to test a great number before finding several which have potential. A jar test is one of the easiest tests to perform but is essentially qualitative. After the conditioned slurry is allowed to settle, the clarity of the supernatant and structure of the floc can be visually observed. Observation of the speed of floc formation, floc size, and settling rate can quickly eliminate those conditioners which have little positive effect. A better test and one which provides quantitative data is the capillary suction time (CST). This test can also be completed with a minimum of time and equipment. Once those polymers which significantly improve filterability have been identified, they can be compared in terms of cost. Since the cost of polymers varies considerably it is advantageous to plot CST versus cost of polymer/weight of dry solids for those polymers under consideration. The two tests outlined above are applicable for initial screening regardless of which dewatering process is involved.

Once the range of polymers has been narrowed down to a few, the effect of various levels of conditioning can be determined. Bench tests

which are specific to a particular dewatering process are normally used to determine the optimum level of conditioning. These will be discussed in relation to the design of equipment in a later section. More general information about the effect of polymer dosage can be generated through particle size analysis and rheology. The performance achieved by a dewatering unit on any sludge is greatly affected by the particle size distribution (percentage of small particles) and the strength of the sludge floc.

As discussed in Section 2.2 the particle size distribution can be determined using a Coulter Counter. As the polymer dosage increases the mean particle diameter increases and the size distribution shifts. For the sludge illustrated in Figure 3-1 the distribution shifted from a concentration of very small particles (<50μ), without conditioning, to one where 83% of the total volume is composed of particles which have a diameter greater than 98μ, at a polymer dosage of 4.5 kg/t.

The yield stress determined from rheograms, as discussed in Section 2.3 gives an indication of the strength of the floc and of how resistant to shear the flocculated sludge is. Yield stress increases as polymer dosage increases for a specific sludge (Figure 3-2). At some level of polymer addition the curve departs from the typical sludge rheogram and a peak is evident at very low shear rates. For any given sludge, the polymer dosage at which this peak is produced is related to the solids concentration of the sludge: the higher the solids concentration the higher the dosage required to produce this peak. Previous work (Campbell and LeClair, 1974) has shown that the polymer dosage at the peak compares favourably with the dosage required for efficient dewatering on pilot plant units. It is suggested that the occurrence of this peak is a particle size phenomenon and may indicate that a certain particle size is one of the critical conditions which must be met to achieve good dewatering.

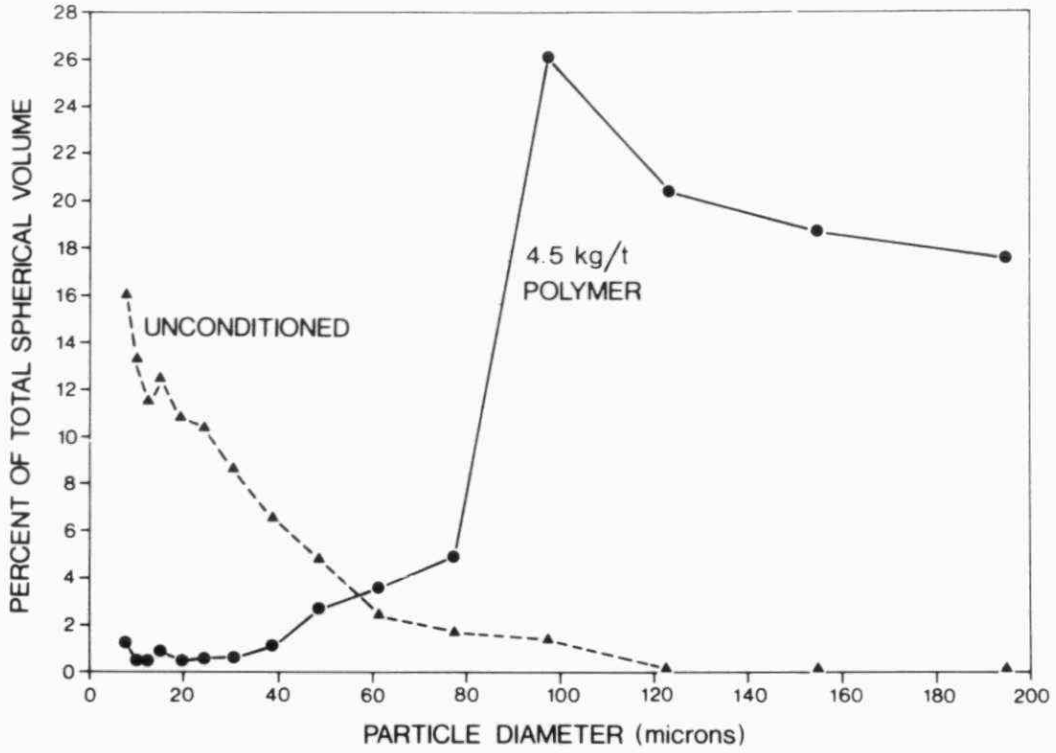


FIGURE 3-1. THE EFFECT OF POLYMER ADDITION ON PARTICLE SIZE DISTRIBUTION

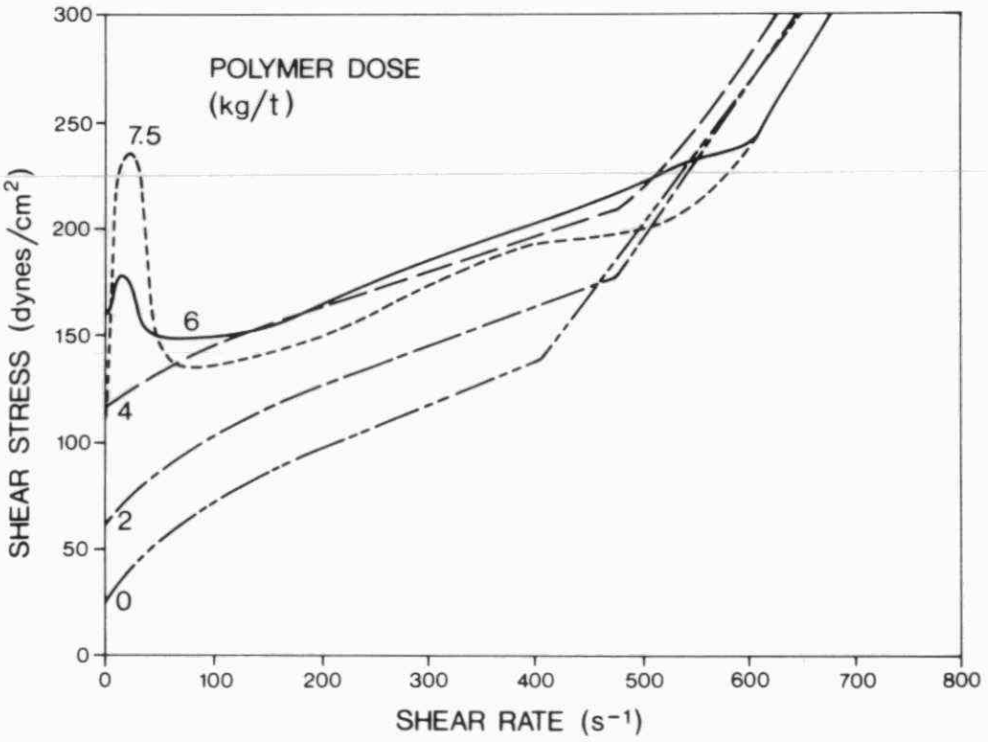


FIGURE 3-2. EFFECT OF POLYMER DOSAGE ON THE RHEOLOGY OF A DIGESTED ALUM SLUDGE

# 4 SLUDGE QUANTITIES

## 4.1 Estimating Sludge Production - Conventional Biological Treatment

The design of sludge handling equipment involves selecting a suitable unit process and estimating the required capacity based on the characteristics and total volume of sludge which must be processed. The relative amounts of primary and secondary sludge will determine the overall sludge characteristics and affect such design parameters as mass loading rates and sludge conditioning requirements. Knowledge of the individual amounts of primary and secondary sludge is also important in evaluating the feasibility of thickening or dewatering either of the sludges separately.

The sludge quantity generated will be a function of the raw wastewater characteristics, the waste treatment process efficiency, and the type of waste treatment process. Methods of estimating solids production during conventional biological treatment have been discussed by several authors including Fair et al (1968); Metcalf and Eddy (1972); Kormanik (1972); and Benefield et al (1975). In addition to the estimating methods based on wastewater characteristics, typical sludge production values reported in the literature will be discussed in the following sections.

### 4.1.1 Primary sludge

Values of primary sludge production from conventional systems, as reported in the literature, are presented in Table 4-1. The data from field surveys in Ontario (Hamoda and Cohen, 1978; Antonic et al, 1978) indicates a lesser volume and mass of sludge produced per unit volume of wastewater treated. This is believed to be a function of the higher degree of thickening achieved in the primary clarifiers (ranges from 3.5% to 8.0%) and the generally dilute nature of Ontario's raw wastewater (i.e., average SS = 173 mg/ℓ, average $BOD_5$ = 122 mg/ℓ). The larger quantity (volume and mass) of sludge produced on a per capita basis is related to the relatively high per capita wastewater flow (660 ℓ/cap-d).

TABLE 4-1. PRIMARY SLUDGE PRODUCTION

| Reference | Sludge Production | | | |
|---|---|---|---|---|
| | Volume Basis | | Mass of Dry Solids Basis | |
| | ℓ/1000 $m^3$ of Wastewater Treated | ℓ/cap-d | kg/1000 $m^3$ of Wastewater Treated | g/cap-d |
| Imhoff and Fair (1940) | 3530 | 1.1 | 179 | 54 |
| Urquart (1972) | 3500 | - | - | - |
| Merritt (1968) | 2500 | - | - | - |
| Imhoff et al (1971) | 3530 | 1.07 | 179 | 54 |
| Metcalf and Eddy (1972) | 2950 | 1.1 | 150 | 54 |
| U.S. EPA (1974) | - | - | 122 | - |
| Knight et al (1973) | - | - | 104 | - |
| Adrian and Smith (1972) | 4465 | - | 95 | - |
| Hamoda and Cohen (1978) | 2000 | 1.32 | 120 | 77 |
| Antonic et al (1978) | 1780 | 1.36 | 89 | 68 |

### 4.1.2 Secondary sludge

Combined primary plus waste activated sludge production values from conventional treatment plants, as reported in the literature, are presented in Table 4-2. The volume of sludge produced per unit volume of sewage treated is significantly lower for the Ontario plants (Hamoda and Cohen, 1978; Antonic et al, 1978) and is due in part to the concentration of the mixed sludge (2% to 7%). Sludge production values for Ontario on any other basis are comparable to the other literature values cited.

TABLE 4-2. COMBINED SLUDGE PRODUCTION (PRIMARY AND WASTE ACTIVATED)

| Reference | Sludge Production | | | |
|---|---|---|---|---|
| | Volume Basis | | Mass of Dry Solids Basis | |
| | ℓ/1000 $m^3$ of Wastewater Treated | ℓ/cap-d | kg/1000 $m^3$ of Wastewater Treated | g/cap-d |
| Imhoff and Fair (1940) | 6180 | 1.87 | 281 | 85 |
| Imhoff et al (1971) | - | 2.0 | - | 90 |
| Metcalf and Eddy (1972) | 6900 | 2.6 | 281 | 106 |
| U.S. EPA (1974) | - | - | 188 | - |
| Knight et al (1973) | - | - | 188 | - |
| Adrian and Smith (1972) | 13,565 | - | 175 | - |
| Wildi (1972) | - | - | - | 91 |
| Hamoda and Cohen (1978) | 3810 | 2.5 | 173 | 114 |
| Antonic et al (1978) | 3880 | 2.6 | 172 | 95 |

### 4.1.3 Digested sludge

Table 4-3 summarizes the reported values of anaerobically digested mixed primary plus activated sludge production from conventional treatment plants. Although data are extremely limited, there are significant differences between the reported values. This illustrates the point that published sludge production values should not be used for design purposes. They are useful for preliminary estimates but each situation must be assessed on its own merits.

TABLE 4-3. ANAEROBIC DIGESTED SLUDGE PRODUCTION (MIXED PRIMARY AND WASTE ACTIVATED)

| Reference | Sludge Production | | | |
|---|---|---|---|---|
| | Volume Basis | | Mass of Dry Solids Basis | |
| | ℓ/1000 $m^3$ of Wastewater Treated | ℓ/cap-d | kg/1000 $m^3$ of Wastewater Treated | g/cap-d |
| Imhoff and Fair (1940) | 2520 | 0.76 | 182 | 55 |
| Imhoff et al (1971) | - | 0.8 | - | 55 |
| Wildi (1972) | - | - | - | 58 |
| Merritt (1968) | 800 | - | - | - |
| Urquart (1972) | 800 | - | - | - |
| Antonic et al (1978) | 1360 | 1.23 | 53 | 41 |

4.1.4 Example calculation of sludge production - conventional biological treatment

As mentioned in Section 4.1, several methods of estimating sludge production are available in the literature. One approach, as outlined below, is based on Figure 4-1, which is a generalized schematic of a secondary wastewater treatment plant, adapted from Kormanik (1972). The design equations based on Figure 4-1 for estimating sludge quantities are as follows:

*Primary Sludge*

$$S_p = kX_o \qquad (4\text{-}1)$$

where: $S_p$ = primary sludge (kg/h).
$k$ = fraction of SS removed in primary clarifier.
$X_o$ = influent SS (kg/h).

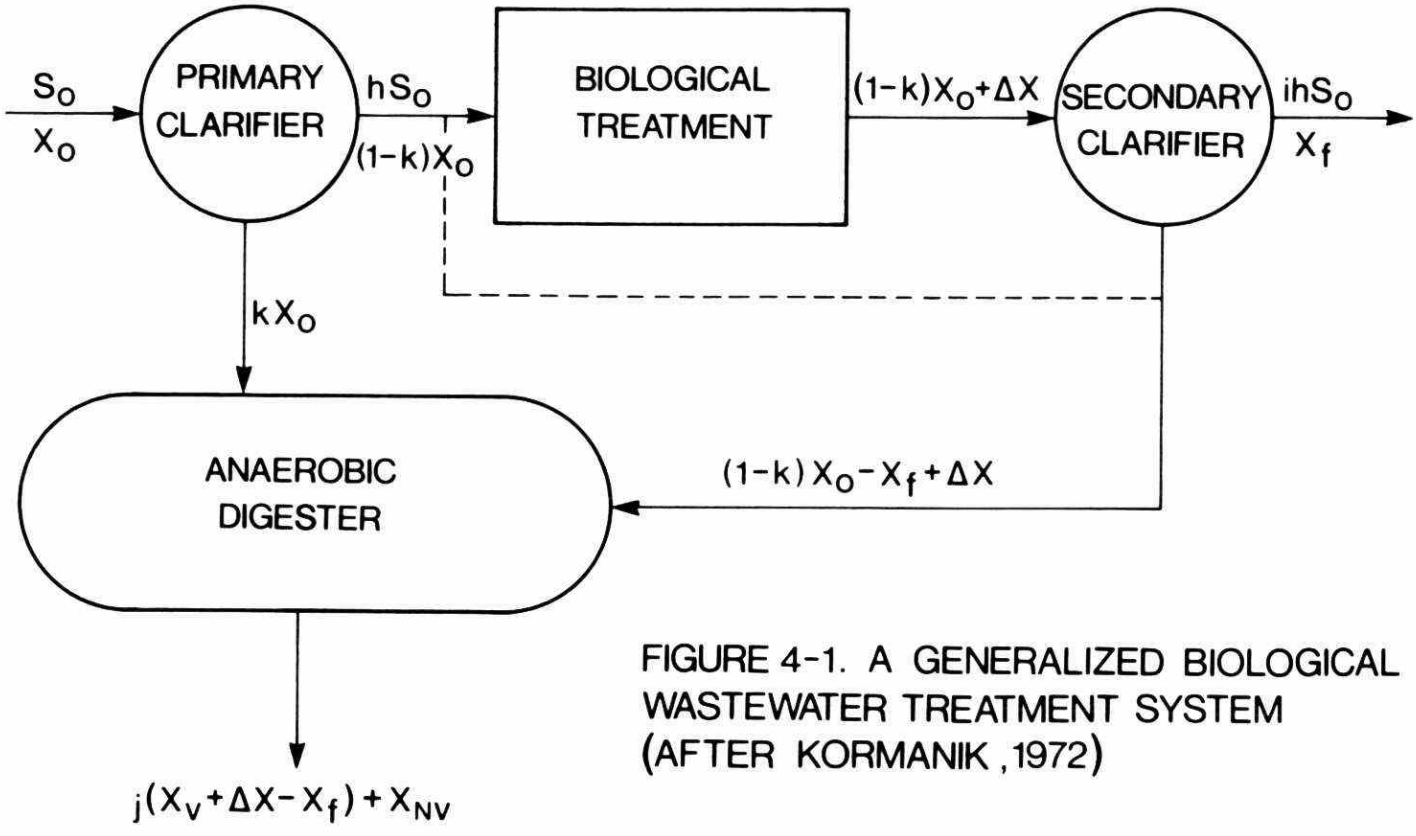


FIGURE 4-1. A GENERALIZED BIOLOGICAL WASTEWATER TREATMENT SYSTEM (AFTER KORMANIK, 1972)

*Secondary Sludge*

$$S_s = (1-k)\ X_o - X_f + \Delta X \qquad (4\text{-}2)$$

where: $S_s$ = secondary sludge (kg/h).
$X_f$ = effluent SS (kg/h).
$\Delta X$ = net solids production by biological action (kg/h).

Net solids production is given by the following equation:

$$\Delta X = (\Delta S)\ Y \qquad (4\text{-}3)$$

where: $\Delta S$ = $BOD_5$ removed by biological action (kg/h).
$Y$ = sludge yield (kg of sludge/kg of $BOD_5$ removed).

$BOD_5$ removal can be calculated according to:

$$\Delta S = h\ S_o - i\ h\ S_o \qquad (4\text{-}4)$$

where: $h$ = fraction of $BOD_5$ remaining after primary clarification.
$S_o$ = influent $BOD_5$ (kg/h).
$i$ = fraction of $BOD_5$ remaining after biological treatment.

*Digested Sludge*

$$S_D = j(X_V + \Delta X - X_f) + X_{NV} \qquad (4\text{-}5)$$

where: $S_D$ = digested sludge (kg/h).
$j$ = fraction of volatile solids remaining after digestion.
$X_V$ = volatile influent SS (kg/h).
$X_{NV}$ = non volatile influent SS (kg/h).

Typical values for raw wastewater to Ontario sewage treatment plants (Antonic et al, 1978) are as follows:

$S_o$ = 122 mg/ℓ
$X_o$ = 173 mg/ℓ
$X_f$ = 15 mg/ℓ
$X_V = 0.7\ X_o$

Values from the literature (Metcalf and Eddy, 1972; Benefield et al, 1975; Fair et al, 1968) for the remaining design variables are as follows:

$k$ = 0.6 (60% SS removal in primary clarifier).
$h$ = 0.7 (30% $BOD_5$ removal in primary clarifier).
$i$ = 0.1 (90% $BOD_5$ removal by biological system).
$j$ = 0.3 (70% VSS removed by anaerobic digestion).
$Y$ = 0.5 (kg of sludge/kg of $BOD_5$ removed).

EXAMPLE PROBLEM

1. Activated sludge treatment plant.
2. Anaerobic digestion.
3. Raw wastewater flow 1000 $m^3/h$.
4. Design values as listed above.

Estimate daily sludge volumes and masses.

*Primary Sludge*

$k = 0.6$

$$X_o = 173 \frac{mg}{\ell} \times \frac{kg}{10^6 mg} \times \frac{10^3\ \ell}{m^3} \times 10^3 \frac{m^3}{h}$$

$$= 173 \text{ kg/h.}$$

Substituting into:

$$S_p = kX_o \qquad (4\text{-}1)$$

$$= 0.6 \times 173 \text{ kg/h} \times 24 \text{ h/d}$$

$$= 2491 \text{ kg/d}$$

Assume the sludge has a solids content of 5% and a specific gravity of 1.02 ($1.02 \times 10^3$ $kg/m^3$).

Daily volume of sludge:

$$\frac{2491 \text{ kg}}{d} \times \frac{kg}{0.05 \text{ kg}} \times \frac{m^3}{1.02 \times 10^3 \text{ kg}} = 48.8 \text{ m}^3/d$$

*Secondary Sludge*

$h = 0.7$

$i = 0.1$

$y = 0.5$ (kg of sludge/kg of $BOD_5$ removed)

$k = 0.6$

$X_o = 173$ kg/h

$$X_f = 15 \frac{mg}{\ell} \times \frac{kg}{10^6 \text{ mg}} \times \frac{10^3\ \ell}{m^3} \times \frac{10^3\ m^3}{h} = 15 \text{ kg/h}$$

$$S_o = 122 \frac{mg}{\ell} \times \frac{kg}{10^6\ mg} \times \frac{10^3\ \ell}{m^3} \times 10^3 \frac{m^3}{h}$$

$$= 122\ kg/h.$$

Substituting into:

$$\Delta S = hS_o - ihS_o \qquad (4\text{-}4)$$

$$= 0.7\ (122) - (0.1)\ (0.7)\ (122)$$

$$= 77\ kg/h.$$

Substituting into:

$$\Delta X = (\Delta S)Y \qquad (4\text{-}3)$$

$$= 77 \times 0.5$$

$$= 38.5\ kg/h.$$

Substituting into:

$$S_s = (1\text{-}k)\ X_o - X_f + \Delta X \qquad (4\text{-}2)$$

$$= (1\text{-}0.6)\ 173 - 15 + 38.5$$

$$= 93\ kg/h \times 24\ h/d$$

$$= 2232\ kg/d.$$

Assume the sludge has a solids content of 1% and a specific gravity of 1.0 ($10^3\ kg/m^3$).

Daily volume of sludge:

$$\frac{2232\ kg}{d} \times \frac{kg}{0.01\ kg} \times \frac{m^3}{10^3\ kg} = 223\ m^3/d$$

*Digested Sludge*

$j = 0.3.$

$X_V = 0.7X_o = 121.1$ kg/h.

$\Delta X = 38.5$ kg/h.

$X_f = 15$ kg/h.

$X_{NV} = 0.3X_o = 51.9$ kg/h.

Substituting into:

$$S_D = j(X_v + \Delta X - X_f) + X_{NV} \qquad (4\text{-}5)$$

$$= 0.3\ (121.1 + 38.5 - 15) + 51.9$$

$$= 95 \text{ kg/h} \times 24 \text{ h/d}$$

$$= 2280 \text{ kg/d}.$$

Assume the sludge has a solids content of 4% and a specific gravity of 1.02 ($1.02 \times 10^3$ kg/m$^3$).
Daily volume of sludge:

$$\frac{2280 \text{ kg}}{\text{d}} \times \frac{\text{kg}}{0.04 \text{ kg}} \times \frac{\text{m}^3}{1.02 \times 10^3 \text{ kg}} = 56 \text{ m}^3/\text{d}.$$

SUMMARY

1. Primary Sludge - 2491 kg/d.
   - 48.8 m$^3$/d.
2. Secondary Sludge - 2232 kg/d.
   - 223 m$^3$/d.
3. Digested Sludge - 2280 kg/d.
   - 56 m$^3$/d.

## 4.2 Estimating Sludge Production - Chemical Precipitation

Chemical clarification of municipal wastewater is now practiced at many locations to remove phosphorus or to improve the overall effluent quality in an overloaded treatment plant. The major inorganic precipitant chemicals

used to date have been alum, ferric chloride, and lime. Addition of these chemicals results in increased sludge production. Sludge dewatering characteristics may also differ considerably from those of a sludge produced in conventional biological treatment.

Chemicals can be added at various points in the treatment sequence. The amount and characteristics of the resulting sludges will depend on the particular point of addition. In general:

1. When chemicals are added before the primary clarifiers, primary sludge production increases (due to the improved clarifier efficiency and the chemical sludge produced), and biological sludge production decreases (due to reduced $BOD_5$ loading to the aeration tanks).
2. When $A\ell^{3+}$ or $Fe^{3+}$ is added to the aeration tanks the increase in solids mass is due almost entirely to the chemical sludge produced.
3. When chemicals are added to the secondary effluent (tertiary treatment) the additional sludge mass is due to increased suspended solids and organics removals, as well as to the increased phosphate and metal hydroxide sludges produced.

A summary of the average values reported in the literature for the amounts of sludge typically produced when $Fe^{3+}$, $A\ell^{3+}$, or lime are added for phosphorus removal at various points in a wastewater treatment system is presented in Table 4-4. The wide range of chemical sludge production values are due to the range of raw wastewater types, chemical dosages and effluent qualities at different locations. The extremely high values reported by Adrian and Smith (1972) for the volume of sludge produced by the addition of either $Fe^{3+}$ or $A\ell^{3+}$ to the primary, are assumed to have been the result of very low sludge concentrations.

Most sanitary engineering textbooks give typical design values for anaerobically digested sludge production. These are very general and should be used with caution. The effect of chemical sludge on solids reduction during digestion was surveyed by Hamoda and Cohen (1978) who found that volatile solids reduction decreased from 45% to 41% as a result of $A\ell^{3+}$ or

TABLE 4-4. CHEMICAL SLUDGE PRODUCTION

| Point of Addition | Chemical | Dosage (mg/ℓ) | Volume ℓ/1000 $m^3$ of wastewater | Mass kg/1000 $m^3$ of wastewater | Reference |
|---|---|---|---|---|---|
| N.S | N.S | N.S | 5120 | 396 | Metcalf and Eddy (1972) |
| N.S | N.S | N.S | 5120 | 258 | Imhoff et al (1971) |
| Primary | $Al^{3+}$ | 13.5 | - | 60 | Vesilind (1973) |
| " | $Al^{3+}$ | 6.3 | - | 228 | Campbell et al (1975) |
| " | $Al^{3+}$ | 7.5 | - | 75 | Stepko and Vachon (1978) |
| " | $Al^{3+}$ | 18.5 | - | 78 | Stepko and Vachon (1978) |
| " | $Al^{3+}$ | N.S | 2380 | 140 | Antonic et al (1978) |
| " | $Al^{3+}$ | 5-18 | 3200 | 170 | Hamoda and Cohen (1978) |
| " | $Al^{3+}$ | 13-23 | 23,000 | 158 | Adrian and Smith (1972) |
| " | $Fe^{3+}$ | 41 | - | 84 | Vesilind (1973) |
| " | $Fe^{3+}$ | 3.0 | - | 261 | Campbell et al (1975) |
| " | $Fe^{3+}$ | 6.9 | - | 100 | Campbell et al (1975) |
| " | $Fe^{3+}$ | 13.6 | - | 161 | Campbell et al (1975) |
| " | $Fe^{3+}$ | 9.6 | - | 80 | Stepko and Vachon (1978) |
| " | $Fe^{3+}$ | 26.6 | - | 92 | Stepko and Vachon (1978) |
| " | $Fe^{3+}$ | N.S | 2380 | 140 | Antonic et al (1978) |
| " | $Fe^{3+}$ | 10-30 | 3200 | 170 | Hamoda and Cohen (1978) |
| " | $Fe^{3+}$ | 26 | 21,920 | 333 | Adrian and Smith (1972) |
| " | $Ca(OH)_2$ | 460 | - | 756 | Vesilind (1973) |
| " | $Ca(OH)_2$ | 150 | - | 253 | Stepko and Vachon (1978) |
| " | $Ca(OH)_2$ | 200 | - | 308 | Stepko and Vachon (1978) |
| " | $Ca(OH)_2$ | 275 | - | 343 | Stepko and Vachon (1978) |
| " | $Ca(OH)_2$ | 100-200 | 3550 | 240 | Hamoda and Cohen (1978) |
| " | $Ca(OH)_2$ | 350-500 | 8925 | 676 | Adrian and Smith (1972) |
| Aeration Tank | $Al^{3+}$ or $Fe^{3+}$ | N.S | 4620 | 200 | Antonic et al (1978) |
| Aeration Tank | $Al^{3+}$ or $Fe^{3+}$ | N.S | 5145 | 217 | Hamoda and Cohen (1978) |
| Aeration Tank | $Al^{3+}$ | 9-23 | 13,480 | 142 | Adrian and Smith (1972) |
| Aeration Tank | $Fe^{3+}$ | 10-30 | 18,650 | 205 | Adrian and Smith (1972) |
| Secondary Effluent | $Al^{3+}$ | 16 | 12,000 | 240 | Adrian and Smith (1972) |
| | $Fe^{3+}$ | 10-30 | 22,070 | 61 | Adrian and Smith (1972) |
| | $Ca(OH)_2$ | 268-450 | 53,400 | 558 | Adrian and Smith (1972) |

N.S. - Not specified.

$Fe^{3+}$ addition for phosphorus removal. Minton and Carlson (1976) estimated that, because of the increased inorganic solids in lime sludge, a total solids reduction in the range of only 20% to 35% would typically result during lime sludge digestion, compared to ≈50% for conventional sludges.

Although the sludge production values reported in the literature are useful for a first approximation, a more precise estimate is required in order to size sludge handling equipment with any degree of confidence. Methods have been developed to provide estimates of sludge quantities based on the stoichiometric reactions involved and the raw and treated wastewater characteristics.

Methods of estimating chemical sludge production on this basis have been given in textbooks (Metcalf and Eddy, 1972; and Fair et al, 1968), phosphorus removal design manuals (U.S. EPA 1971; U.S. EPA 1976; Knight et al, 1973) and other recent literature (Stepko and Vachon, 1978; Farrell, 1975; Minton and Carlson, 1976). By far the most detailed discussion of estimating chemical sludge production has been given by Farrell (1975) who also contributed to the revised EPA Phosphorus Removal Design Manual (U.S. EPA, 1976).

Because of space limitations in this report the reader is referred to either Farrell (1975) or EPA (1976) for details. A relatively brief discussion and examples of chemical sludge production estimating procedures are presented below under the appropriate headings.

### 4.2.1 Example calculation of sludge production - chemical precipitation

*Alum Addition*

The $Al^{3+}$ requirements are based on the following equations:

$$\underset{(27)}{Al^{3+}} + \underset{(95 \text{ or } 31 \text{ as } P)}{PO_4^{3-}} \rightarrow \underset{(122)}{AlPO_4\downarrow} \qquad (4\text{-}6)$$

$$\underset{(27)}{Al^{3+}} + \underset{(51)}{3(OH)^-} \rightarrow \underset{(78)}{Al(OH)_3\downarrow} \qquad (4\text{-}7)$$

Typical $A\ell^{3+}$ dosages required to achieve ≃ 1 mg/ℓ residual total P are reported to range from 5 to 18 mg/ℓ $A\ell^{3+}$, depending upon the initial wastewater conditions (Hamoda and Cohen, 1978; Wei and LeClair, 1975). Dosage requirements can also be predicted on the basis of Aℓ/P ratios. Ratios between 2.0 and 3.1 have been used and recommended by several authors including Farrell (1975) and Prested et al (1977).

*Iron Addition*

The $Fe^{3+}$ requirements are based on the following equations:

$$\underset{(56)}{Fe^{3+}} + \underset{(95 \text{ or } 31 \text{ as P})}{PO_4^{3-}} \rightarrow \underset{(151)}{FePO_4\downarrow} \qquad (4\text{-}8)$$

$$\underset{(56)}{Fe^{3+}} + \underset{(51)}{3(OH)^-} \rightarrow \underset{(107)}{Fe(OH)_3\downarrow} \qquad (4\text{-}9)$$

Typical $Fe^{3+}$ dosages required to achieve ≃ 1 mg/ℓ residual total P are reported to range from 10 to 30 mg/ℓ $Fe^{3+}$, depending upon the initial wastewater conditions (Hamoda and Cohen, 1978; Wei and LeClair, 1975; Prested et al 1977). On the basis of Fe/P ratios, Wei and LeClair (1975) used a range of 1.71 to 3.81, Prested et al (1977) found the average to be 2.5, and Farrell (1975) recommended a ratio of 2.2.

*Lime Addition*

The lime requirements are based on the following equations:

$$\underset{(370)}{5Ca(OH)_2} + \underset{(288 \text{ or } 93 \text{ as P})}{3HPO_4^{2-}} \rightarrow \underset{(502)}{Ca_5OH(PO_4)_3\downarrow} + \underset{(102)}{6OH^-} + \underset{(54)}{3H_2O} \qquad (4\text{-}10)$$

$$\underset{(24)}{Mg^{2+}} + \underset{(34)}{2OH^-} \rightarrow \underset{(58)}{Mg(OH)_2\downarrow} \qquad (4\text{-}11)$$

$$\underset{(74)}{Ca(OH)_2} + \underset{(61)}{HCO_3^-} \rightarrow \underset{(100)}{CaCO_3\downarrow} + \underset{(18)}{H_2O} + \underset{(17)}{OH^-} \qquad (4\text{-}12)$$

The pH required to achieve a given level of phosphorus removal is very dependent on the initial raw wastewater hardness and other characteristics. Minton and Carlson (1976) give an excellent discussion of this relationship. $Mg(OH)_2$ starts to precipitate above pH≈10.5 and in general, the higher the hardness and $Mg^{2+}$ concentration the lower the pH required to achieve a given phosphorus effluent quality (Prested et al, 1977).

Once the required pH has been established, the lime dosage can be estimated on the basis of the alkalinity of the raw wastewater (Minton and Carlson, 1976; Farrell, 1975).

EXAMPLE PROBLEM

Calculate the mass of chemical sludge solids produced during chemical addition to the primary using 1) alum, 2) ferric chloride, and 3) lime under the following conditions:

| Influent | | Effluent from the Primary | |
|---|---|---|---|
| $BOD_5$ | = 200 mg/ℓ. | $BOD_5$ | = 70 mg/ℓ (65% removal). |
| SS | = 250 mg/ℓ. | SS | = 38 mg/ℓ (85% removal). |
| $PO_4^{3-}$ | = 6.3 mg/ℓ as P. | $PO_4^{3-}$ | = 0.9 mg/ℓ (85% removal) as P. |
| $Ca^{2+}$ | = 60 mg/ℓ. | | |
| $Mg^{2+}$ | = 20 mg/ℓ. | | |
| Alkalinity | = 275 mg/ℓ as $CaCO_3$. | | |

1. *Alum Addition*

Assume a dosage of 13 mg/ℓ $Al^{3+}$ (Al/P = 2.1) to achieve a 0.9 mg/ℓ P residual (Farrell, 1975).

Phosphorus removed = 6.3 - 0.9 = 5.4 mg/ℓ.

From Equation 4-6:

31 mg of P produces 122 mg of $AlPO_4$

$$5.4 \text{ mg/}\ell \text{ P removed} \times \frac{122 \text{ mg } A\ell PO_4}{31 \text{ mg P}} = 21.3 \text{ mg/}\ell$$

Solids production due to P removal = 21.3 mg/ℓ $A\ell PO_4$

From Equation 4-6, $A\ell^{3+}$ used in $A\ell PO_4$ formation is:

$$5.4 \text{ mg/}\ell \text{ P} \times \frac{27 \text{ mg } A\ell^{3+}}{31 \text{ mg P}} = 4.7 \text{ mg/}\ell \text{ } A\ell^{3+}$$

Excess $A\ell^{3+}$ = 13 - 4.7 = 8.3 mg/ℓ.

Excess $A\ell^{3+}$ goes to $A\ell(OH)_3$ according to Equation 4-7:

$$8.3 \text{ mg/}\ell \text{ } A\ell^{3+} \times \frac{78 \text{ mg } A\ell(OH)_3}{27 \text{ mg } A\ell^{3+}} = 24 \text{ mg/}\ell$$

Solids production due to $A\ell(OH_3)$ formation = 24.0 mg/ℓ $A\ell(OH)_3$

Total solids production [$A\ell PO_4$ + $A\ell(OH)_3$] ≃ 45 mg/ℓ

2. *Iron Addition*

Assume a dosage of 15 mg/ℓ $Fe^{3+}$ (Fe/P = 2.38) to achieve a 0.9 mg/ℓ P residual.

From Equation 4-8:

31 mg of P produces 151 mg of $FePO_4$

$$5.4 \text{ mg/}\ell \text{ of P removed} \times \frac{151 \text{ mg } FePO_4}{31 \text{ mg P}} = 26.3 \text{ mg/}\ell$$

Solids production due to P removal = 26.3 mg/ℓ $FePO_4$

From Equation 4-8, $Fe^{3+}$ used in $FePO_4$ formation is:

$$5.4 \text{ mg/}\ell \text{ P} \times \frac{56 \text{ mg } Fe^{3+}}{31 \text{ mg P}} = 9.8 \text{ mg/}\ell$$

Excess $Fe^{3+}$ (15 - 9.8 = 5.2) goes to $Fe(OH)_3$ according to Equation 4-9:

$$5.2 \text{ mg/}\ell \ Fe^{3+} \times \frac{107 \text{ mg } Fe(OH)_3}{56 \text{ mg } Fe^{3+}} = 9.9 \text{ mg/}\ell$$

Solids production due to $Fe(OH)_3$ formation = 9.9 mg/ℓ $Fe(OH)_3$

Total solids production [$FePO_4$ + $Fe(OH_3)$] ≃ 36 mg/ℓ

3. *Lime Addition*

A pH of 11.0 was selected to achieve a 0.9 mg/ℓ P residual (≃85% P removal) (Rush et al, 1976).

Based on the alkalinity of 275 mg/ℓ as $CaCO_3$, a dosage of 400 mg/ℓ $Ca(OH)_2$ will be required to raise the pH to 11.0 (Farrel, 1975).

Assume $Ca(OH)_2$ is 95% active.

Required lime dosage = 400/0.95 = 421 mg/ℓ.

Solids production due to inerts in lime is:

421 mg/ℓ x 0.05 = 21.1 mg/ℓ

From Equation 4-10:

93 mg of P produces 502 mg $Ca_5OH(PO_4)_3$

$$5.4 \text{ mg of P removed} \times \frac{502 \text{ mg } Ca_5OH(PO_4)_3}{93 \text{ mg P}} = 29.1 \text{ mg/}\ell$$

Solids production due to P removal = 29.1 mg/ℓ

Assume that soluble $Mg^{2+}$ in the effluent = 2.0 mg/ℓ at a pH of 11.0 (Rush et al, 1976).

$Mg^{2+}$ removed = 20 - 2 = 18 mg/ℓ.

From Equation 4-11:

$$18 \text{ mg/ℓ } Mg^{2+} \text{ removed} \times \frac{58 \text{ mg } Mg(OH)_2}{24 \text{ mg } Mg^{2+}} = 43.5 \text{ mg/ℓ}$$

Solids production due to $Mg(OH)_2$ formation = 43.5 mg/ℓ

From Equation 4-10, $Ca(OH)_2$ consumed in phosphorus removal is:

$$5.4 \text{ mg/ℓ P} \times \frac{370 \text{ mg } Ca(OH)_2}{93 \text{ mg P}} = 21.5 \text{ mg/ℓ}$$

This is equivalent to 11.3 mg $Ca^{2+}$ [in $Ca_5OH(PO_4)_3$]

$Ca^{2+}$ In = $Ca^{2+}$ in raw wastewater + $Ca^{2+}$ from $Ca(OH)_2$

$$= 60 \text{ mg/ℓ} + 400 \text{ mg/ℓ } Ca(OH_2) \times \frac{40 \text{ mg } Ca^{2+}}{74 \text{ mg } Ca(OH)_2}$$

$$= 276 \text{ mg/ℓ } Ca^{2+} = Ca^{2+} \text{ Out}$$

At a pH of 11.0 soluble $Ca^{2+}$ in the effluent = 60 mg/ℓ (Farrel, 1975)

$Ca^{2+}$ Out = $Ca^{2+}$ soluble in effluent + $Ca^{2+}$ in $Ca_5OH(PO_4)_3$ + excess $Ca^{2+}$ (precipitated as $CaCO_3$)

$$= 60 \text{ mg/ℓ} + 11 \text{ mg/ℓ} + \text{excess } Ca^{2+}$$

Excess $Ca^{2+}$ = 276 - 71 = 205 mg/ℓ

According to Equation 4-12:

40 mg $Ca^{2+}$ [74 mg $Ca(OH)_2$] produces 100 mg $CaCO_3$

$$205 \text{ mg/}\ell \text{ } Ca^{2+} \times \frac{100 \text{ mg } CaCO_3}{40 \text{ mg } Ca^{2+}} = 512.5 \text{ mg/}\ell$$

Solids production due to $CaCO_3$ formation = 512.5 mg/ℓ

Total solids production [Lime Inerts + $Ca_5OH(PO_4)_3$ + $Mg(OH)_2$ + $CaCO_3$] = 606 mg/ℓ

The $CO_3$ present in the alkalinity (275 mg/ℓ) would produce 550 mg/ℓ of $CaCO_3$; therefore there is sufficient for the calculated value of 512.5 mg/ℓ.

Additional $CaCO_3$ solids may be generated if effluent recarbonation is practiced (U.S. EPA 1976; Farrell, 1975).

A summary of the sludge production estimates for the example is presented in Table 4-5.

TABLE 4-5. SLUDGE PRODUCTION SUMMARY - CHEMICAL ADDITION TO THE PRIMARY CLARIFIER (BASED ON STOICHIOMETRY)

| Chemical / Sludge Production | Alum | Iron | Lime |
|---|---|---|---|
| kg/1000 $m^3$ of wastewater treated | 45 | 36 | 606 |
| Primary Sludge (Section 4.1.4) kg/1000 $m^3$ of wastewater treated | 103 | 103 | 103 |
| Total Primary Sludge kg/1000 $m^3$ of wastewater treated | 148 | 139 | 709 |

Other factors which affect total sludge production during chemical addition include:

1. When addition is to the primary clarifier, suspended solids removal will increase to 85% to 95%, $BOD_5$ removal will increase to ≃75%, and ≃30% of the soluble organics may be removed (Farrell, 1975). The latter two effects will increase primary sludge production and decrease biological sludge production.
2. When addition is to the aeration tank, increased solids production is almost entirely due to the chemical solids, and therefore results in no increase of SS removal (Farrell, 1975).
3. Sludge volume to be handled is a function of the solids concentration achieved and will depend on the degree of thickening which can be achieved.
4. More chemical sludge is produced than predicted from the stoichiometry (Farrell, 1975; Sutton et al, 1977). This is thought to be due mainly to the incorporation of other metal cations into the complex ligand structures of the aluminum, ferric and magnesium hydroxides precipitated. Consequently, the U.S. EPA (1976) recommends adding 35% to the estimate of solids production during $A\ell^{3+}$ or $Fe^{3+}$ addition and 20% to the estimate during lime addition. More research is required to expand the existing data base.

## 4.3 Effect of Phosphorus Removal on Sludge Production and Dewatering

A general discussion of chemical solids production, including estimating procedures was presented in the previous section. The overall effect of this additional sludge on the design of sludge handling equipment will vary with each waste treatment plant, and depends on many variables including sludge solids concentration, increase in either volume or mass and dewaterability of the combined sludge.

Recent pilot scale and full scale experience with sludges from installations practicing chemical precipitation of phosphorus is summarized as follows:

Stepko and Vachon (1978) carried out phosphorus removal demonstration studies at CFB Borden (primary treatment only) using lime, alum and ferric chloride addition to the primary clarifier. Solids concentrations increased from 3.8% to an average of 13.5% during lime addition. Alum and ferric chloride addition increased the solids concentration marginally from 3.8% to an average of 4.5%. The increase in sludge mass over baseline values ranged from 495% to 784% for lime, 31% to 72% for alum, and 58% to 78% for ferric chloride. Although the increase in solids concentration tended to minimize the effect, the net result in the sludge volume was a 100% increase with lime addition, a 14% to 27% increase with alum, and 14% to 53% with ferric chloride. The dewatering properties of these chemical sludges were tested on a rotary drum belt vacuum filter with dimensions of 0.914 m x 0.305 m (3 ft x 1 ft). Maximum yields for lime and $FeC\ell_3$ sludges were 40.2 kg/m$^2$-h (8.3 lb/ft$^2$-h) and 22.4 kg/m$^2$-h (4.6 lb/ft$^2$-h), respectively. Dewatering of the alum sludge was not successful due to poor cake discharge characteristics. Dewatering data was not reported for the baseline sludge.

Hamoda and Cohen (1978) compared sludge production at Ontario treatment plants with no chemical addition, to those adding chemicals for phosphorus removal. Experience at eight upgraded primary plants using alum or ferric salts showed an average increase of 60% and 40% for volume and mass of sludge produced, respectively. The addition of lime to one primary plant resulted in an increase of mass of 200%, from 80 to 250 kg/1000 m$^3$ of wastewater. The solids content rose from 5.7% to 8% and the volume increased 57%, from 2260 to 3550 $\ell$/1000 m$^3$ of wastewater. Based on mean values, the total sludge production increase (primary plus waste activated sludge) at upgraded activated sludge plants with chemical addition to the aeration tank, was 35% and 26% for volume and mass, respectively. The weighted average of sludge solids decreased from 4.5% to 4.2%. One secondary plant adding lime to the primary had a 156% increase in dry solids mass but only a 21% increase in volume due to an increase in the total solids concentration from 3.8% to 8%.

Full scale chemical addition for phosphorus removal has been carried out at several Canadian treatment plants for a number of years and has been reported on by various authors (Van Fleet et al, 1972; Boyko and Rupke, 1973; Campbell et al, 1975; Hamoda and Cohen, 1978). Dry solids production (mass) increased by 5% to 25% in waste activated sludge plants using metal salts. In primary plants there was a 100% increase in sludge production because of increase in solids capture. In the secondary plants average solids concentration dropped from 4.75% to 4.25%, while in the primary plants sludge solids concentration decreased by 20%.

Records of two plants which had practiced full scale sludge dewatering both before and after the introduction of phosphorus removal were surveyed to determine the effects of implementing phosphorus removal (Campbell et al, 1975). In 1975 the use of both alum and ferric chloride was evaluated at the West Windsor Sewage Treatment Plant (primary treatment only). A summary of the data generated during the test periods, along with baseline data from 1973, is shown in Table 4-6. Sludge production increased significantly with chemical addition but part of the increase could be attributed to an increase in the suspended solids loading to the plant. The concentration of the primary sludge, the filter yield and the filter cake solids decreased, while the filtrate solids, and conditioning costs increased. These effects were more pronounced with alum addition than with ferric chloride and resulted in the selection of ferric chloride as the precipitant for phosphorus removal. The North Toronto Sewage Treatment Plant has been using ferric chloride for phosphorus removal since 1973. A summary of their experience is also presented in Table 4-6. The major consequences of adding ferric chloride were an increase in the sludge conditioning requirements and a decrease in the filter cake solids. Significant changes were not evident in the sludge production or filter yield.

The wide range of values reported in this section outline dramatically the need for an accurate estimate of sludge quantities. The use of generalized sludge production values for design is not recommended, and could result in either inadequate sludge handling equipment or unneeded excess capacity. Where possible, sludge estimates should be based on a knowledge of the characteristics of the raw wastewater, the unit processes within the treatment sequence, and the records of similar plants already in operation.

TABLE 4-6. FULL SCALE VACUUM FILTRATION OF SLUDGES FROM PHOSPHORUS REMOVAL FACILITIES

| | West Windsor[2] | | | North Toronto[2] | |
|---|---|---|---|---|---|
| Raw Sewage Flow ($m^3/h$) | 4940 | 3960 | 4070 | 1600 | 1570 |
| Phosphorus Removal Chemical | None | $Fe^{3+}$ | $Al^{3+}$ | None | $Fe^{3+}$ |
| Phosphorus Removal Cost[1] ($/1000 $m^3$) | 0 | 6.20 | 6.32 | 0 | 2.80 |
| Raw Sewage SS (mg/ℓ) | 110 | 164 | 195 | 206 | 193 |
| Sludge Production (kg/1000 $m^3$ of raw sewage) | 93 | 161 | 228 | 97 | 100 |
| Type of Sludge | Primary | Primary | Primary | Digested Elutriated | Digested Elutriated |
| Solids Concentration (%) | 11.9 | 8.6 | 7.9 | 8.1 | 7.6 |
| Conditioning Chemicals (% Lime) (% Ferric Chloride) | 9.9 1.3 | 15.9 0.1 | 24.0 1.2 | 9.5 0 | 11.8 6.6 |
| Conditioning Cost[1] ($/tonne of dry solids) | 4.41 | 7.36 | 8.13 | Not Recorded | Not Recorded |
| Filter Yield ($kg/m^2$ - h) | 60.5 | 46.8 | 32.7 | 18.5 | 19.0 |
| Filter Cake Solids (%) | 31 | 21 | 17 | 23 | 19 |
| Filtrate SS (mg/ℓ) | 2830 | 3660 | 13900 | 5550 | 7690 |

[1]Cost Figures as of December, 1975.
[2]Campbell et al (1975).

# 5 DESIGN METHODOLOGY

The design of a sludge dewatering process can be based on either experimental data generated for the sludge in question or on experience with similar sludges. In situations where sludge is not available for testing, such as the design of the dewatering section of a complete new treatment plant, there is little choice but to base the design on past experience. Many dewatering installations, however, constitute either an additional phase of the treatment complex (e.g., phased construction) or an expansion of existing dewatering facilities (e.g., due to plant expansion or modification). Under these circumstances sludge is available and the design approach should be based on experimental data. The following section outlines bench and pilot plant tests applicable for gravity thickening, dissolved air flotation, centrifugation, vacuum filtration, and pressure filtration, and how they are used in the design procedure. The development of theoretical equations will be dealt with only briefly since complete derivations are available in the literature.

The use of bench and pilot plant tests to generate design data will be stressed. Equipment manufacturers who are required to guarantee the performance of a dewatering unit are often reluctant to quote on a contract without carrying out a pilot plant test program first, even though bench tests may be available for that particular process. The use of bench tests in a situation such as this can be a decided advantage in reducing the number of pilot scale experiments required. If the chemical conditioner and the range of dosages required can be selected, and the effect of some of the operating variables established at the bench scale level, then a maximum of information can be obtained at the pilot plant level with a minimum input of time and cost.

The data used to illustrate the design procedure for each process are based on experimental results generated at the Wastewater Technology Centre in Burlington, Ontario, as part of sludge dewatering studies carried out under the research and development program of the Canada-Ontario Agreement on Great Lakes Water Quality. All assumptions regarding the full scale flow rates and required performance levels are presented with each design example. Two aspects of design which are not considered in the

examples are the optimum operating schedule and the need for reserve capacity during equipment breakdown or overhaul. These factors have a significant effect on the size of equipment specified and should be based on the requirements of each individual installation. It is common practice, in plants with a design capacity of 45 450 $m^3/d$ (10 mgd) or greater, to operate the sludge dewatering equipment on a continuous, seven days a week basis. It has been assumed, for the purpose of illustrating design techniques only, that all equipment will operate 24 hours a day, seven days a week, without reserve capacity. Although it is obvious that this assumption is not realistic for smaller plants in particular, it is a simple matter to adjust the hourly flow rate to compensate for the decreased operating schedule.

Complete data for a process as outlined in the design procedure was not always available for the particular sludge in question. In these instances typical data for experiments on other sludges is presented and identified as being for illustration purposes only.

## 5.1 Gravity Thickening

### 5.1.1 Process description and theory

Gravity thickening is often used as a preliminary low cost method of sludge volume reduction prior to digestion or dewatering. Gravity thickeners are essentially conventional clarifiers that have been modified to accommodate a greater volume of sludge, have heavier raking mechanisms and may have pickets attached to the rakes.

The design of thickeners is based on the assumption that most sewage sludges settle as a blanket in the zone settling regime. The rate of subsidence of this blanket, for a given sludge, will be a constant which is characteristic of the concentration. As concentration layers build up from the bottom they will eventually intersect the sludge liquid interface, and the settling velocity will ultimately decrease to zero. The objective of thickener design is to ensure that the rate of application of solids does not exceed the rate at which solids are transmitted to the bottom.

A gravity thickener can be operated in either a batch or continuous mode. In a batch thickener solids are transported to the thickener bottom

only by the influence of gravity. The capacity for carrying solids downward, referred to as the batch flux, is given by the following equation:

$$G_B = C_i V_i \qquad (5\text{-}1)$$

where: $G_B$ = batch solids flux ($kg/m^2$-h).
$C_i$ = feed sludge concentration ($kg/m^3$).
$V_i$ = settling velocity at $C_i$ (m/h).

In a continuous thickener a net downward force is also exerted by the removal of thickened sludge from the bottom. The solid flux due to this effect is:

$$G_u = C_i U \qquad (5\text{-}2)$$

where: $G_u$ = flux due to sludge removal ($kg/m^2$-h).
$U$ = downward velocity due to sludge removal (m/h).

The value of U is given by:

$$U = Q_u/A \qquad (5\text{-}3)$$

where: $Q_u$ = volumetric flow rate of sludge removal ($m^3/h$).
$A$ = thickener area ($m^2$).

The total flux ($G_T$) in a continuous thickener is given by adding equations (5-1) and (5-2).

$$G_T = G_B + G_u$$

or:
$$G_T = C_i V_i + C_i U \qquad (5\text{-}4)$$

In order to specify the area of a thickener it is necessary to determine the limiting flux ($G_L$) under which a thickener will operate for a given set

of operating conditions. Figure 5-1 illustrates the use of a typical batch flux plot to determine the limiting flux ($G_L = G_T$) for a continuous thickener. Detailed development of thickener design theory is available in the literature (Dick, 1970 and Vesilind, 1974c).

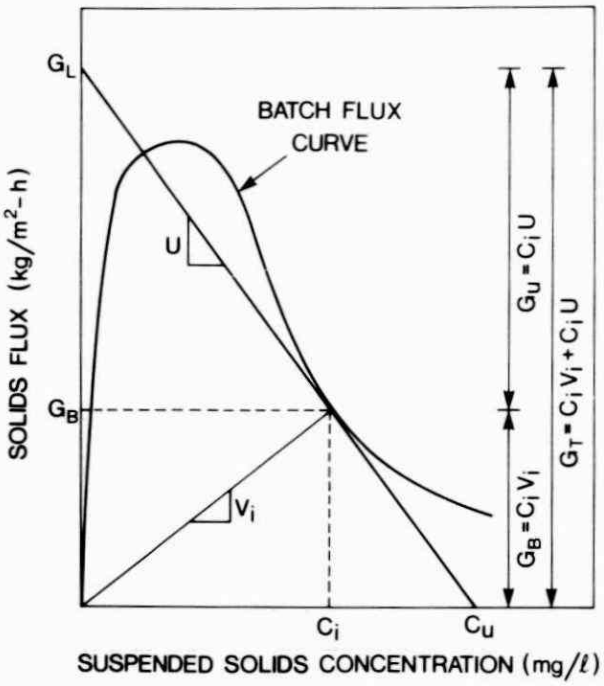


FIGURE 5-1. DETERMINATION OF LIMITING FLUX FOR A CONTINUOUS THICKENER USING A BATCH FLUX PLOT (REFERENCE: DICK (1970))

The performance of a gravity thickener is affected by the *solids loading rate*, *underflow rate*, *detention time*, *rake speed*, and *sludge characteristics*. The first group of parameters (*loading*, *underflow*) are interrelated through the solids flux curve. A change in one produces a corresponding change in the other. An increase in the volume available for sludge storage will increase the *detention time* and should increase the concentration of the sludge. If the detention time is excessive, gas formation may occur and result in the flotation of some of the solids. If the *rake speed* is too fast, redispersion of the settled solids may occur. Changes in the *sludge characteristics* due to chemical

conditioning can enhance performance. Polymers can result in higher underflow concentrations but are more often used to improve the quality of the effluent.

### 5.1.2 Development of design information

Design data can be generated by either batch or continuous tests. In practice, a continuous flow bench-scale thickener is seldom used due to the problems of maintaining steady state conditions. The small scale of the equipment, the hoses, and the pumping requirements for both feed and underflow do not readily lend themselves to handling thickened sludge. Such systems are often plagued with stoppages in the lines, overflowing of the thickener and generally erratic operation.

The most common method of obtaining design data is to conduct a series of settling tests at several different influent solids concentrations. In order to minimize the inadequacies of laboratory settling tests, Vesilind (1974c) has suggested the following procedure:

1. The cylinder should be at least 20 cm in diameter.
2. The initial height should be the same as for the prototype thickener. When this is not practical, 90 to 100 cm should be considered minimum.
3. The cylinder should be filled from the bottom.
4. The sludge should be slowly stirred throughout the test (0.5 rpm for a 20 cm cylinder).

The height of the interface is plotted against elapsed time for each test to produce a series of settling curves (Figure 5-2). A complete set of 12 settling tests were conducted with initial solids concentrations ranging from 0.04% to 1.8%, but for clarity the results of only five tests are shown in Figure 5-2. The zone settling velocity determined from the initial linear portion of each curve is plotted against the sludge solids concentration on semi-log graph paper (Figure 5-3). The settling velocity as determined from the best fit line (method of least squares) in Figure 5-3, multiplied by the initial concentration, gives a value for the batch flux. A plot of batch flux versus solids concentration results in a batch flux curve (Figure 5-4).

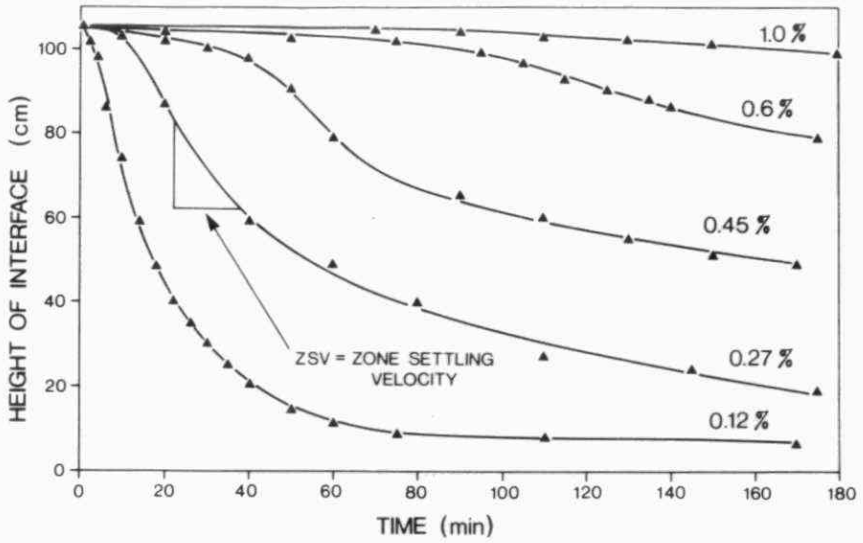


FIGURE 5-2. WASTE ACTIVATED SLUDGE SETTLING CURVES

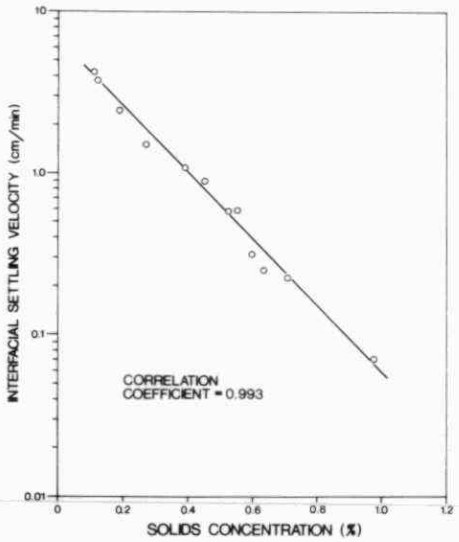


FIGURE 5-3. INTERFACIAL SETTLING VELOCITY VERSUS SOLIDS CONCENTRATION

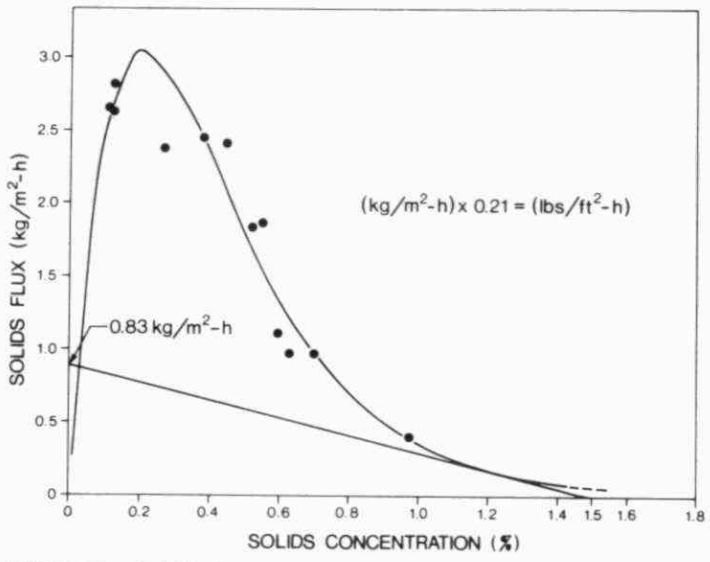


FIGURE 5-4. BATCH FLUX CURVE FOR WASTE ACTIVATED SLUDGE

The limiting solids flux is determined by drawing a straight line from the desired underflow concentration and tangent to the batch flux curve. The point of intersection of this line with the vertical axis gives the value of the limiting solids flux. Selection of the required surface area is based upon the following relationship (Adams and Eckenfelder, 1974):

$$A = Q_o C_o / G_L \qquad (5\text{-}5)$$

where: $A$ = surface area ($m^2$).
$Q_o$ = influent flow rate ($m^3/h$).
$C_o$ = influent solids concentration ($kg/m^3$).
$G_L$ = limiting solids flux ($kg/m^2\text{-}h$).

The flow rate of the thickened sludge underflow can be determined on the basis of the following mass balance around the thickener (Adams and Eckenfelder, 1974):

$$Q_o C_o = Q_u C_u + C_e (Q_o - Q_u) \qquad (5\text{-}6)$$

where: $Q_u$ = underflow ($m^3/h$).
$C_u$ = underflow solids concentration ($kg/m^3$).
$C_e$ = supernatant suspended solids ($kg/m^3$).

The designer should be aware of the limitations of the procedure outlined above. When a straight line does not adequately represent the data as presented in Figure 5-3, a different approach is used to construct the batch flux curve. The zone settling velocity as determined in Figure 5-2 is multiplied by the appropriate initial solids concentration to obtain a single batch flux value for use in constructing the batch flux curve (Figure 5-4). Under these circumstances a substantially larger number of settling tests are normally required to produce a smooth batch flux curve.

5.1.3 Design example

SLUDGE CHARACTERISTICS

1. Waste activated sludge.
2. Flow rate = 25 $m^3/h$ (90 Igpm).
3. Influent solids concentration = 0.55%.

PERFORMANCE CRITERIA

1. Underflow concentration = 1.5%.
2. Supernatant suspended solids < 100 mg/ℓ.

A series of settling tests without the addition of polymer produced the results shown in Figures 5-2, 5-3, and 5-4. In all cases, the suspended solids of the supernatant were less than 100 mg/ℓ. The limiting solids flux for an underflow concentration of 1.5% is determined from Figure 5-4. Alternative designs could be obtained by selecting different underflow concentrations.

Calculation of Surface Area:

$Q_o$ = 25 $m^3/h$.
$C_o$ = 0.55% = 5.5 $kg/m^3$.
$G_L$ = 0.83 $kg/m^2$-h.

Substituting into:

$$A = Q_o C_o / G_L \qquad (5\text{-}5)$$

$$= \frac{25 \times 5.5}{0.83} = 166 \text{ m}^2 \ (1786 \text{ ft}^2)$$

The required thickener diameter would be 14.54 m (47.7 ft). This would normally be supplied by the next largest standard diameter. Based on experience, the depth of gravity thickeners has generally been standardized at 3 - 4 m (10 - 13 ft).

Calculation of Underflow Rate:

$Q_o$ = 25 $m^3/h$.
$C_o$ = 0.55% = 5.5 $kg/m^3$.
$C_u$ = 1.5% = 15.0 $kg/m^3$.
$C_e$ = 100 mg/ℓ = 0.1 $kg/m^3$.

Substituting into:

$$Q_o C_o = Q_u C_u + C_e (Q_o - Q_u) \quad (5\text{-}6)$$

$$25 \times 5.5 = Q_u 15 + 0.1 (25 - Q_u)$$

$$Q_u = 9.1 \text{ m}^3/\text{h} \text{ (33 Igpm)}$$

Pumping capacity of 9.1 $m^3/h$ (33 Igpm) would be required for sludge withdrawal.
If a higher level of performance was necessary (e.g., $C_u$ = 4.0%) with the same sludge, the effect of polymer addition could be evaluated. This would consist of repeating the settling tests at the appropriate polymer dosages and developing a new batch flux curve for each dosage.

DESIGN SUMMARY

| | |
|---|---|
| Flow Rate | 25 $m^3/h$ (90 Igpm). |
| Initial Solids Concentration | 0.55%. |
| Sludge Quantity | 1204 tonne/year. |
| Solids Loading (flux) | 0.83 $kg/m^2$-h(0.17 $lbs/ft^2$-h). |
| Polymer Addition | None. |
| Underflow Concentration | 1.5%. |
| Underflow Rate | 9.1 $m^3/h$ (33 Igpm). |
| Supernatant Suspended Solids | <100 mg/ℓ. |
| Surface Area | 166 $m^2$ (1786 $ft^2$). |
| Thickener Diameter | 14.54 m (47.7 ft). |
| Thickener Depth | 3 - 4 m (10 - 13 ft). |

## 5.2 Dissolved Air Flotation

### 5.2.1 Process description and theory

Dissolved air flotation is widely accepted as a practical technique for separating oils and greases from a liquid, and as a means of thickening activated sludge and flocculant chemical sludges. By attaching air bubbles to the suspended particles until their combined net specific gravity becomes less than that of water, the particles are made buoyant. These buoyant particles then rise quickly to the surface forming a "float" which can be removed by skimming. To produce the microscopic air bubbles required for effective flotation, air is added to a liquid (water) and the mixture elevated to a high pressure [276 to 621 $kN/m^2$ (40 to 90 psig)]. Since the solubility of air increases with pressure, the air dissolves. When the pressure is released, the liquid becomes supersaturated and the volume of air in excess of the atmospheric saturation concentration comes out of solution in the form of microscopic bubbles. The air required for flotation can be provided by pressurization of either the total incoming sludge flow or a recycled flow of clarified effluent. In sewage treatment applications recycle pressurization is the more common since the fragile sludge flocs tend to be destroyed when passed through a pressurization system.

The quantity of air which will be released from solution when the pressure is reduced to atmospheric levels is given by the following equation:

$$A^* = R\, C_s\, (fP - 1) \qquad (5\text{-}7)$$

where: $A^*$ = quantity of air released (kg/h).

$R$ = pressurized recycle flow ($m^3/h$).

$C_s$ = gas saturation at atmospheric conditions ($kg/m^3$)**.

$f$ = fraction of saturation of air dissolved in the pressurization system.

$P$ = pressure in the retention tank (atmospheres).

---

** $mg/\ell \times 10^3 = kg/m^3$.

The quantity of solids which must be floated is given by:

$$S = Q X_o \tag{5-8}$$

where: $S$ = solids flow rate (kg/h).
$Q$ = influent flow rate ($m^3/h$).
$X_o$ = influent suspended solids ($kg/m^3$).

The ratio of the weight of air released to the weight of solids present ($A^*/S$) has traditionally been used as the primary design parameter for flotation. The air/solids ratio can be calculated by the following equation:

$$\frac{A^*}{S} = \frac{R\ C_s\ (fP - 1)}{Q\ X_o} \tag{5-9}$$

A good description of the theoretical aspects of flotation is available in the literature (Ettelt, 1964).

The process variables which affect the performance of a flotation unit include *saturation pressure*, *recycle rate*, *air to solids ratio*, *solids loading*, *hydraulic loading*, and *chemical conditioning*. *Saturation pressures* on most commercial units range from 276 to 414 $kN/m^2$ (40 to 60 psig). At lower pressures it is difficult to get enough air dissolved in the recycle flow while higher pressures tend to result in the production of larger bubbles which are less effective (Wood and Dick, 1973). The *recycle rate* determines the amount of air available for flotation under a fixed pressure and retention tank design. Except as it relates to the amount of air, the recycle rate only becomes important at levels which create excessive horizontal velocities. An increase in the *air to solids ratio* has generally been reported to enhance both clarification and thickening (Vesilind, 1974c; Eckenfelder and Ford, 1970). A recent study (Gehr, 1975) showed no definite correlation between air/solids ratio and performance over the range of air/solids ratios tested. There is a definite minimum value of air/solids ratio below which flotation will not occur. The maximum *solids loading* to the unit is usually governed by the capacity of the recycle system to

provide air for flotation. Thickening applications are usually limited by the solids loading because of the high suspended solids concentrations in the influent. The *hydraulic loading* is an important factor when the flotation system is used for clarification but in thickening it is seldom a problem. The use of *chemical conditioning* with sewage sludges is usually necessary for stable operation of the flotation system. The addition of polymer increases the suspended solids recovery and the concentration of the float.

### 5.2.2 Development of design information

Information for the design of dissolved air flotation (DAF) units can be generated by either bench scale or pilot scale tests. Bench scale experiments utilize either a two or three cylinder batch flotation apparatus. The procedures for these two tests are described by Eckenfelder and Ford (1970), and by Wood and Dick (1973), respectively. The two systems differ primarily in the method of introducing the pressurized recycle to the sludge, but with sludges amenable to flotation, results are comparable (Wood and Dick, 1973). After the recycle liquid is blended with the sludge, the rise rate of the interface is measured and the test allowed to stand for 20 minutes. Samples of the float and the clarified liquid are then taken for solids analysis.

The batch test is an effective method of evaluating the effect of air/solids ratio and polymer dosage on the rise rates, the float solids concentration, and the effluent suspended solids. Since test conditions do not duplicate continuous operation, application of these data should be limited to a qualitative prediction of performance. Thus, the use of optimum air/solids ratio as determined by the batch test, coupled with empirical values for surface overflow rates (Eckenfelder and Ford, 1970) is not recommended for designing air flotation units for thickening applications.

Theoretically it is possible to use the rise rates measured during batch tests to generate a flux curve in a similar manner to that for gravity thickening (Vesilind, 1974c; Dick, 1970). This approach is still unproven, thus there is no method of evaluating, at the bench scale level, the effect of solids loading on DAF performance.

The float from a batch unit is usually less concentrated than that from a continuous flow unit. In a continuous flow unit, the top of the float is pushed up out of the liquid such that there is a vertical concentration gradient through the float layer. The skimming system is designed to collect only that portion on the top which is most concentrated. In a batch operation there is neither the sustained upward force nor the required depth of float to achieve the same results. When sampling the float on a batch test, care must be taken to avoid collecting water below the interface. Under these circumstances the size of the sample may not be sufficiently large to be representative.

In order for batch test results to adequately predict the performance of continuous units it is essential that the degree of saturation in the recycle liquid be the same for both units. In the past this has seldom been measured but has been assumed to range from 90% to 95%. It has been reported by Gehr (1975) that some full scale units may only achieve levels of 40% to 50%.

The design of a full scale flotation unit involves determination of the required surface area. This is generally based on the solids loading to the unit. The maximum solids loading at which the desired performance can be achieved will be affected by the air/solids ratio and the polymer dosage. In order to determine this maximum loading for any given set of conditions, it is usually necessary to operate a continuous flow pilot plant. The most common pilot plants have surface areas ranging from 0.09 to 1.4 $m^2$ (1 to 15 $ft^2$). Previous studies have shown that comparable performance could be achieved on three sizes of DAF units, 0.09 $m^2$, 1.4 $m^2$ and 37.2 $m^2$ operating on the same sludge (Campbell and LeClair, 1974).

### 5.2.3 Design example

SLUDGE CHARACTERISTICS

1. Waste activated sludge.
2. Flow rate = 25 $m^3/h$ (90 Igpm).
3. Influent solids concentration = 1% = 10 $kg/m^3$.
4. Solids flow rate = 25 $m^3/h$ x 10 $kg/m^3$
   = 250 kg of dry solids/h.

PERFORMANCE CRITERIA

1. Float solids = 4%.
2. Suspended solids recovery $\geq$ 99%.

For sludge conditioning a cationic polymer was chosen on the basis of previous laboratory work.

Batch flotation tests using the two cylinder apparatus produced the results shown in Figures 5-5 and 5-6. Increasing polymer dosage increased the rise rate of the float (Figure 5-5). The effect of polymer dosage on float solids and effluent solids was inconsistent (Figure 5-6) but suspended solids recovery exceeded 99% for all tests.

A 0.09 $m^2$ DAF pilot plant was operated to determine the effect of solids loading, polymer dosage, and air/solids ratio on the performance of the unit. An effluent with less than 30 mg/$\ell$ of suspended solids was achieved under all operating conditions. The effect of the operating variables on the float solids is shown in Figure 5-7.

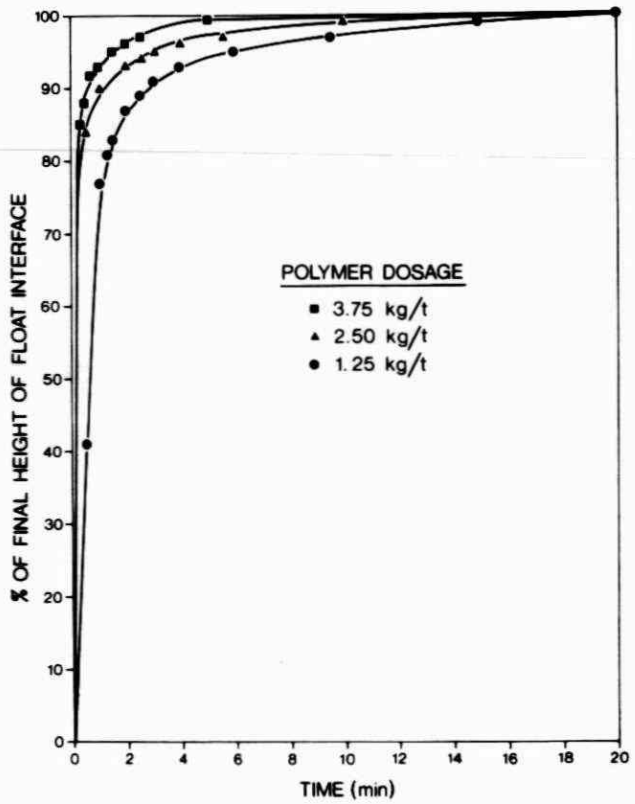


FIGURE 5-5. RISE RATE CURVES

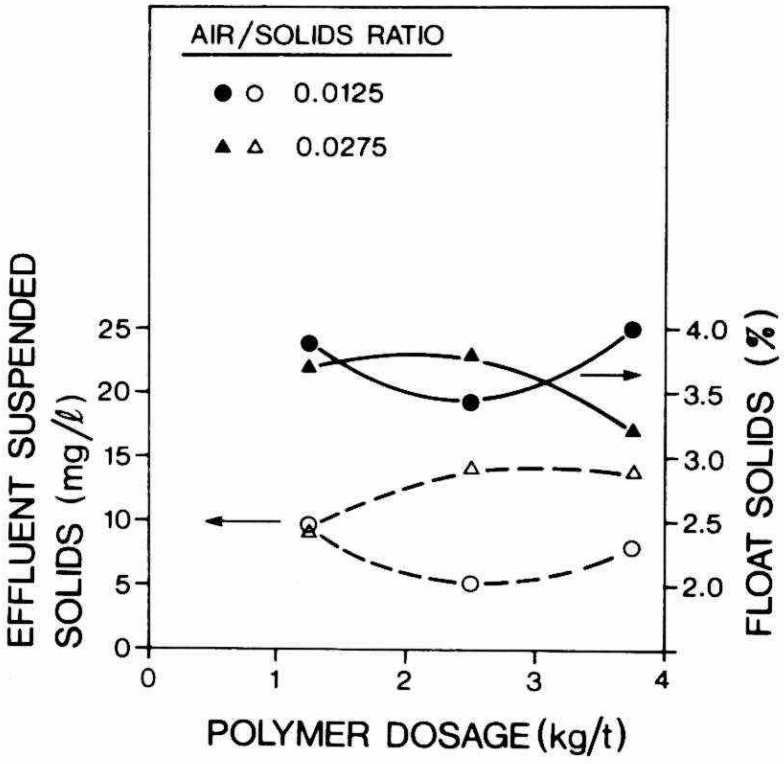


FIGURE 5-6. BATCH AIR FLOTATION TESTS

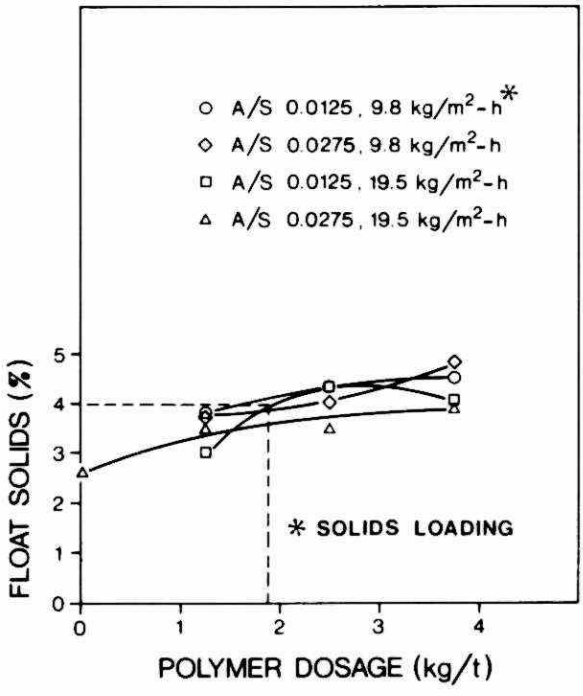


FIGURE 5-7. THICKENING PERFORMANCE OF DISSOLVED AIR FLOTATION PILOT PLANT

The operating variables should be selected to maximize solids loading and minimize polymer dose and air/solids ratio. A solids recovery of 99% would be satisfied under all conditions tested. Figure 5-7 indicates that the required float solids of 4% could be most economically obtained under the following conditions:

Solids loading = 19.5 kg/$m^2$-h (4 lbs/$ft^2$-h).
Polymer dosage = 1.85 kg/t (3.7 lbs/ton).
Air/solids ratio = 0.0125.

Reference to Figure 5-7 indicates that costs could be reduced by operating without polymer addition. The cost reduction achieved would be offset by lower float solids concentrations (2.6% compared to 4%) and by a decrease in the operating stability of the system.

Based on solids loading the required surface area is:

$$\frac{250 \text{ kg/h}}{19.5 \text{ kg/m}^2\text{-h}} = 12.8 \text{ m}^2 \text{ (137 ft}^2\text{)}$$

Since DAF units are normally prefabricated in standard sizes, a unit having an effective surface area of 14 $m^2$ (150 $ft^2$) would be selected.

The effective solids loading would be decreased to:

$$\frac{250 \text{ kg/h}}{14 \text{ m}^2} = 17.9 \text{ kg/m}^2\text{-h (3.66 lbs/ft}^2\text{-h)}$$

This solids loading falls within the suggested operating range of 9.8 to 19.5 kg/$m^2$-h (2 to 4 lbs/$ft^2$-h) (U.S. EPA, 1974).

The hydraulic loading to the unit would be:

$$\frac{25 \text{ m}^3\text{/h}}{14 \text{ m}^2} = 1.8 \text{ m}^3\text{/h -m}^2 \text{ (0.6 lgpm/ft}^2\text{)}$$

This is below the suggested maximum overflow rate for thickening applications of 2.4 $m^3/h\text{-}m^2$ (0.8 Igpm/$ft^2$) (U.S. EPA, 1974).

DESIGN SUMMARY

| | |
|---|---|
| Flow Rate | 25 $m^3/h$ (90 Igpm). |
| Influent Solids Concentration | 1%. |
| Solids Flow Rate | 250 kg/h (550 lbs/h). |
| Sludge Quantity | 2190 tonne/year. |
| Solids Loading | 17.9 $kg/m^2\text{-}h$ (3.66 $lbs/ft^2\text{-}h$). |
| Polymer Dosage | 1.85 kg/t (3.7 lbs/ton). |
| Air/Solids Ratio | 0.0125. |
| Hydraulic Loading | 1.8 $m^3/h\text{-}m^2$ (0.6 Igpm/$ft^2$). |
| Surface Area | 14 $m^2$ (150 $ft^2$). |
| Float Solids | 4%. |
| Effluent Suspended Solids | 30 mg/ℓ. |

## 5.3 Centrifugation

### 5.3.1 Process description and theory

The separation of a liquid-solid slurry in a centrifuge is analagous to the separation process in a gravity sedimentation basin. The basic difference is that in a gravity settler the only applied force is the force of gravity whereas in a centrifuge the applied centrifugal force is 1000 to 3000 times that of gravity. The increased settling velocity imparted by the centrifugal force and the very small distances, particles are required to settle account for the comparatively high liquid capacities of centrifuges. The three types of centrifuges which have found varying degrees of application in the wastewater field are the *solid bowl scroll*, the *imperforate bowl basket* and the *disc-nozzle*.

The *solid bowl scroll* type (Figure 5-8) has been the most popular in the wastewater field. It has the highest solids handling capacity of the three and can also dewater the greatest variety of sludges. The centrifuge consists of a cylindrical-conical bowl with an internal scroll conveyor. Feed sludge is introduced at the centre of the spinning bowl where the particles come under the influence of centrifugal force. Under this force

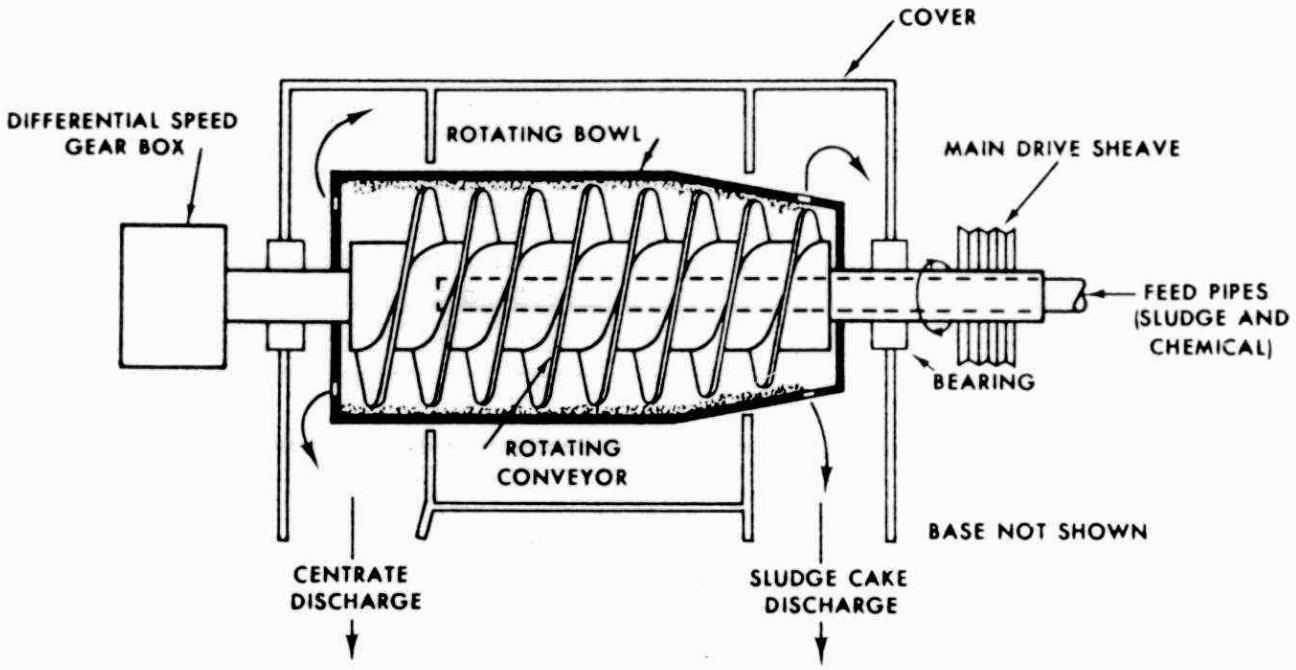


FIGURE 5-8. SOLID BOWL SCROLL CENTRIFUGE
(REFERENCE : U.S. EPA, 1974)

the solids settle outwards forming a layer at the bowl wall. The liquid (centrate) is discharged either by gravity over a weir or removed by a skimmer. The solids are moved by the scroll conveyor along the cylindrical bowl wall and up the conical section (beach) where they are discharged by gravity. The variables which influence centrifuge performance include *feed rate*, *polymer addition*, *bowl speed*, *pool depth* and *conveyor speed*. Table 5-1 shows the effect of each of these parameters on the performance. Solid bowl centrifuges have been applied to all sludge dewatering aspects of municipal waste treatment. These include thickening of waste activated sludge and dewatering of primary, digested or a combination of both.

TABLE 5-1. EFFECT OF OPERATING VARIABLES ON THE PERFORMANCE OF A SOLID BOWL CENTRIFUGE

| Operating Parameter | Performance | |
|---|---|---|
| | Solids Recovery | Cake Solids |
| Increase Feed Rate | Decrease | Increase |
| Increase Polymer Addition | Increase | Variable |
| Increase Bowl Speed[1] | Increase | Increase |
| Increase Pool Depth | Increase | Decrease |
| Increase Conveyor Speed | Decrease | Increase |

[1]The exception occurs with delicate flocculated sludges such as waste activated sludge. In this case an intermediate or lower bowl speed is usually optimum.

The *imperforate bowl basket* centrifuge (Figure 5-9) employs a rotating vertical bowl with a lip ring. Feed is introduced at the centre of the base and as it flows upward the solids settle out into a cake at the bowl wall. The clarified effluent overflows the lip ring at the top and discharges by gravity. No provision is made for sludge cake removal during operation. When the sludge cake has accumulated to the edge of the lip ring or when centrate quality has deteriorated below a predetermined level, the feed is interrupted and the softer liquid interior surface of the cake skimmed off while the bowl is still rotating at full speed. The bowl speed is then reduced to approximately half-speed and the remaining compacted solids are cut out using a stationary advancing knife. The bottom of the bowl opens simultaneously to allow discharge of the solids. After operating speed has been resumed, sludge flow is started and another cycle initiated. The cycles are completely automated and are based either on a simple timer or on the results of a turbidimeter in the effluent line. The variables which affect performance include *sludge feed rate, bowl speed* and *polymer addition* (Table 5-2).

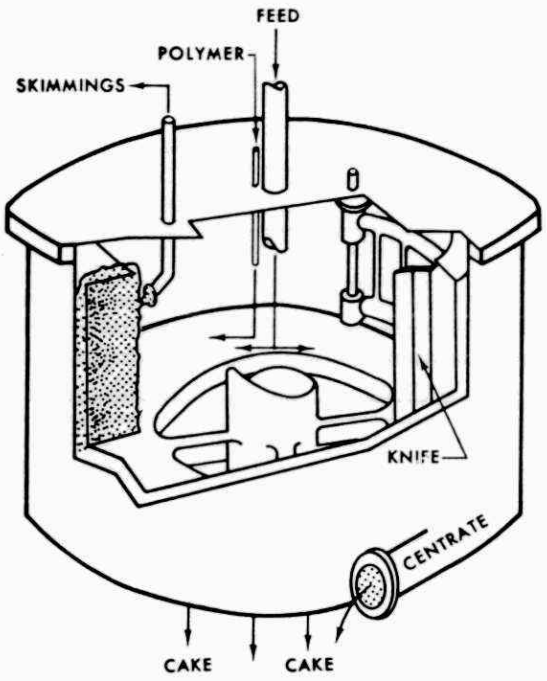


FIGURE 5-9. SCHEMATIC DIAGRAM OF AN IMPERFORATE BOWL BASKET CENTRIFUGE (REFERENCE: U.S. EPA, 1974)

TABLE 5-2. EFFECT OF OPERATING VARIABLES ON THE PERFORMANCE OF AN IMPERFORATE BOWL BASKET CENTRIFUGE

| Operating Parameter | Performance | | |
|---|---|---|---|
| | Solids Recovery | Cake Solids | Cycle Time |
| Increase Feed Rate | Decrease | Decrease | Decrease |
| Increase Bowl Speed | Increase | Increase | Increase |
| Increase Polymer Dosage | Increase | Variable | Increase |

Basket centrifuges are suitable for a wide range of sludges but are most effective when used to dewater light fluffy sludges such as waste activated sludge. Their greatest advantage is that sedimentation occurs in a quiescent zone, undisturbed by sludge removal. This results in high recoveries without the use of polymers. Although the system is automated, operation must be on a batch basis, which results in a relatively low solids handling capacity.

The *disc-nozzle* centrifuge (Figure 5-10) derives its name from the configuration of the separation chamber. The feed is introduced at the centre of the bowl and is distributed through spaces between stacked conical discs. As the feed enters these narrow channels, suspended solids settle radially outward under the influence of centrifugal force. The solids collect on the underside of the discs and eventually slide down the discs and out to the bowl wall where they compact before being discharged through nozzles in the bowl periphery. The clarified centrate emerges from the inner edges of the cones and is discharged separately. The disc-nozzle centrifuge is best suited to removing low concentrations of fine particles from large volumes of wastewater. The only application of disc machines to sludge handling has been the thickening of waste activated sludge. Since even this application has been extremely limited, and experimental data was not available, the disc centrifuge will not be discussed any further in this report.

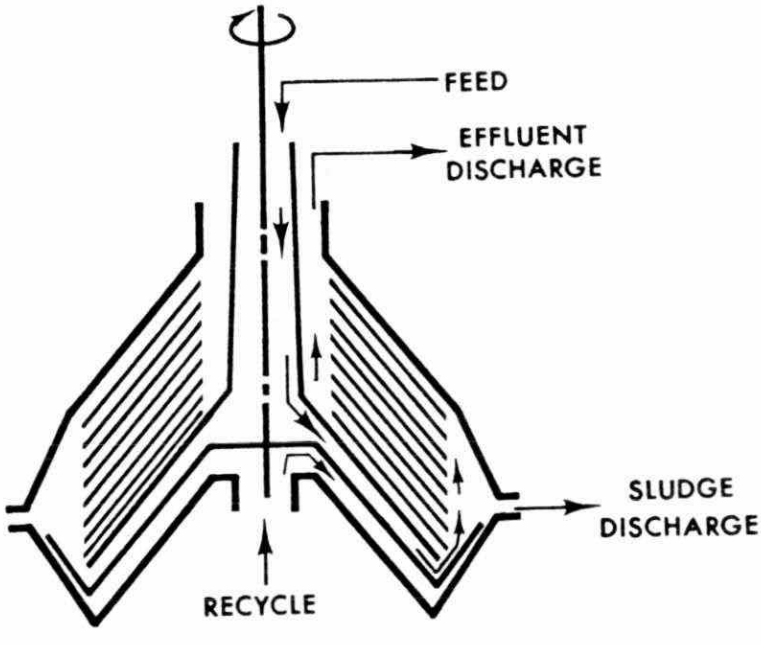


FIGURE 5-10. DISC TYPE CENTRIFUGE
(REFERENCE: U.S. EPA, 1974)

The theoretical hydraulic capacity factor referred to as sigma ($\Sigma$), represents the area of a gravity settling tank which would be required to give equivalent sedimentation performance. The value of sigma for either a solid bowl or imperforate bowl basket centrifuge can be calculated according to the following equation (Lavanchy and Keith, 1967):

$$\Sigma = 2\pi \ell \frac{\omega}{g} (0.75 r_2^2 + 0.25 r_1^2) \quad (5\text{-}10)$$

where: $\ell$ = effective clarifying length of the centrifuge bowl (cm).
$\omega$ = angular velocity of centrifuge bowl (radians/s).
$g$ = acceleration due to gravity (cm/s$^2$).
$r_1$ = radius from centrifuge centreline to the liquid surface in the centrifuge bowl (cm).
$r_2$ = radius from centrifuge centreline to the inside wall of the centrifuge bowl (cm).

A simplified method of calculating $\Sigma$ which is only applicable to a solid bowl centrifuge is given by the following equation (Vesilind, 1974c):

$$\Sigma = \frac{V \omega^2}{g \ell n \frac{r_2}{r_1}} \quad (5\text{-}11)$$

where: $V$ = centrifuge pool volume (cm$^3$).

All other symbols are as listed for equation 5-10.

The theoretical solids handling capacity (m$^3$/h) of a solid bowl centrifuge, referred to as beta ($\beta$), can be determined from the following relationship (Vesilind, 1974c):

$$\beta = \pi \Delta\omega D S N Z \quad (5\text{-}12)$$

where: $\Delta\omega$ = differential speed between scroll and bowl (rph).
$D$ = bowl diameter (m).
$S$ = scroll blade pitch (m).
$N$ = number of leads on the scroll.
$Z$ = pool depth (m).

### 5.3.2 Development of design information

A variety of centrifuge bench tests have been reported in the literature, but have not been used for design purposes. The primary uses of bench tests have been to evaluate qualitatively the feasibility of centrifugation, and to select the type and dosage of polymer required.

Tests which can provide useful information but which are not recommended for design include:

1. Shaking a sample of conditioned sludge in a bottle and observing the stability of the floc (Schultz, 1967).
2. Analysis of sludge samples spun on a lab centrifuge. Settling properties are measured by determining the suspended solids in the centrate while scrolling properties are determined by a penetration test. Correlation of results by simple regression analysis to full scale installations produced an equation for the prediction of solids recovery (Vesilind, 1969).
3. Monitoring the interfacial subsidence rate using a lab centrifuge and strobe light in order to develop a batch flux curve (Vesilind, 1974a).
4. Prediction of maximum cake solids based on a mass balance of samples spun on a lab centrifuge (Vesilind, 1973).

A test developed at the Wastewater Technology Centre may have potential, both as a design tool and as a polymer selection aid. In this test 50 mℓ samples of digested lime sludge were spun on a bench centrifuge at 3500 rpm (≈1000 x G) for two minutes. A portion of the sludge cake was scrolled out of the tube using an auger drill bit while the tube was held at a 45° angle. Solids analyses were done immediately on the liquid remaining in the tube before the dispersed solids had a chance to resettle. One measure of the cake solids concentration was determined by solids analysis on the scrolled portion, while another value was obtained on the basis of the cake volume before scrolling (Vesilind, 1973) and the two values averaged. Correlation between this bench test and a 15 x 30 cm (6 x 12 in) pilot scale solid bowl centrifuge is shown in Figure 5-11. It is anticipated that the effect of flow rate could be reproduced by altering the spin time on the bench

centrifuge. The use of the test as a method for determining suitable polymers and estimating required dosages is illustrated in Figure 5-12. Assuming the costs of the two polymers are equal, polymer A is superior to polymer B in terms of both solids recovery and cake concentration. In the course of future experimental work at the Wastewater Technology Centre, it is intended to further refine this bench test.

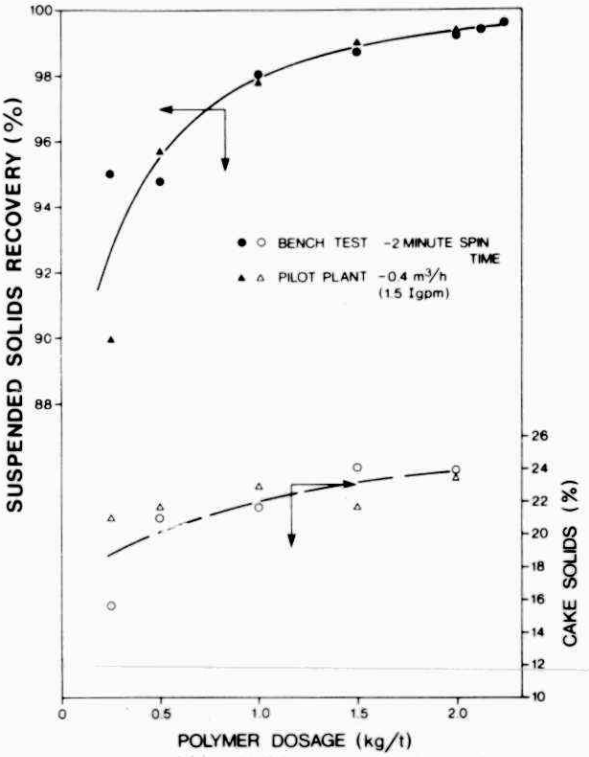


FIGURE 5-11. COMPARISON OF BENCH TEST PREDICTION AND PILOT PLANT PERFORMANCE OF A SOLID BOWL CENTRIFUGE

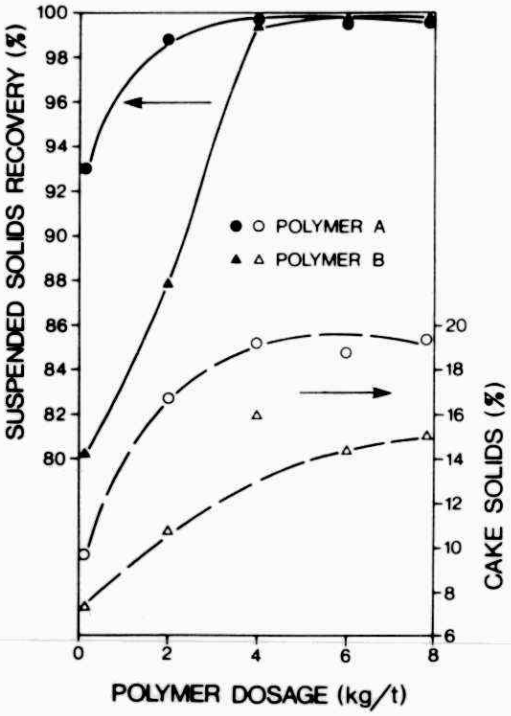


FIGURE 5-12. BENCH TEST PREDICTION OF SOLID BOWL CENTRIFUGE PERFORMANCE

At this time operation of a pilot plant represents the only means of generating reliable data for the design of a full scale *solid bowl centrifuge*. Scale-up of the liquid handling capacity is based on the assumption that when the ratio of hydraulic flow rate ($Q_H$) to the theoretical hydraulic capacity factor ($\Sigma$), as determined by Equation 5-11, is the same for two geometrically similar but different sized machines, then

the performance of the two machines will be similar for a given slurry (Vesilind, 1974b). The scale-up relationship is as follows:

$$\frac{Q_{HP}}{\Sigma_p} = \frac{Q_{HF}}{\Sigma_F} \qquad (5\text{-}13)$$

where: $Q_{HP}$ = hydraulic flow rate ($m^3/h$) of the pilot scale unit

$Q_{HF}$ = hydraulic flow rate ($m^3/h$) of the full scale unit.

$\Sigma_p$ = hydraulic capacity factor ($cm^2$) of the pilot scale unit

$\Sigma_F$ = hydraulic capacity factor ($cm^2$) of the full scale unit.

Scale-up of the solids handling capacity is based on the assumption that when the ratio of solids flow rate ($Q_S$) to the theoretical solids handling capacity ($\beta$), as determined by Equation 5-12, is the same for two geometrically similar but different sized machines, then the performance of the two machines will be similar for a given slurry (Vesilind, 1974b). The scale-up relationship is as follows:

$$\frac{Q_{SP}}{\beta_p} = \frac{Q_{SF}}{\beta_F} \qquad (5\text{-}14)$$

where: $Q_{SP}$ = solids flow rate (kg/h) to the pilot scale unit

$Q_{SF}$ = solids flow rate (kg/h) to the full scale unit.

$\beta_p$ = solids handling capacity ($m^3/h$) of the pilot scale unit

$\beta_F$ = solids handling capacity $m^3/h$) of the full scale unit.

The best design procedure is to develop full scale requirements on the basis of both hydraulic and solids loading considerations. The limiting design would govern machine selection. In most wastewater applications, the centrifuge systems are governed by hydraulic capacity and a design based on sigma values is sufficient. The exception occurs when very concentrated

feed sludges are encountered. In this case, the design may be limited by the solids handling capacity and scale-up should be conducted using both methods.

Design of an *imperforate bowl basket centrifuge* should also be based on pilot plant results. Sigma values can be calculated for the basket centrifuge and a scale-up procedure similar to that outlined for the solid bowl centrifuge used. The sigma concept is based on the assumption of ideal flow conditions but the flow within a basket centrifuge does not satisfy this criterion. As a result manufacturers have, through experience, developed correction factors to compensate for the non-ideal conditions present. The use of sigma to scale-up basket centrifuges, without access to the correction factors, is not recommended. A good approximation of the full scale (liquid and solid) capacity required can be developed based on sludge densities (wet and dry) and centrifuge bowl volumes.

### 5.3.3 Design example

(a) *Solid Bowl Scroll Centrifuge*

SLUDGE CHARACTERISTICS

1. Primary alum sludge.
2. Flow rate = 25 $m^3/h$ (90 Igpm).
3. Feed solids concentration = 6%.

PERFORMANCE CRITERIA

1. Cake solids = 16%.
2. Solids recovery = 99%.

Although not available for this sludge, polymer selection would normally be based on bench test results similar to those illustrated in Figure 5-12. A series of experimental runs on a 15.2 x 30.5 cm (6 x 12 in) pilot scale centrifuge produced the results shown in Figure 5-13. Due to the characteristic light fluffy nature of alum flocs, a deep pool (1.6 cm) and a relatively low centrifugal force (1360 x G or 4000 rpm) were pre-selected

before testing. The performance criteria could be satisfied (Figure 5-13), under the following operating conditions:

Flow rate = 0.68 $m^3/h$ (2.5 Igpm).
Scroll differential speed = 10.5 rpm.
Polymer dosage = 2.5 kg/t (5.0 lbs/ton).

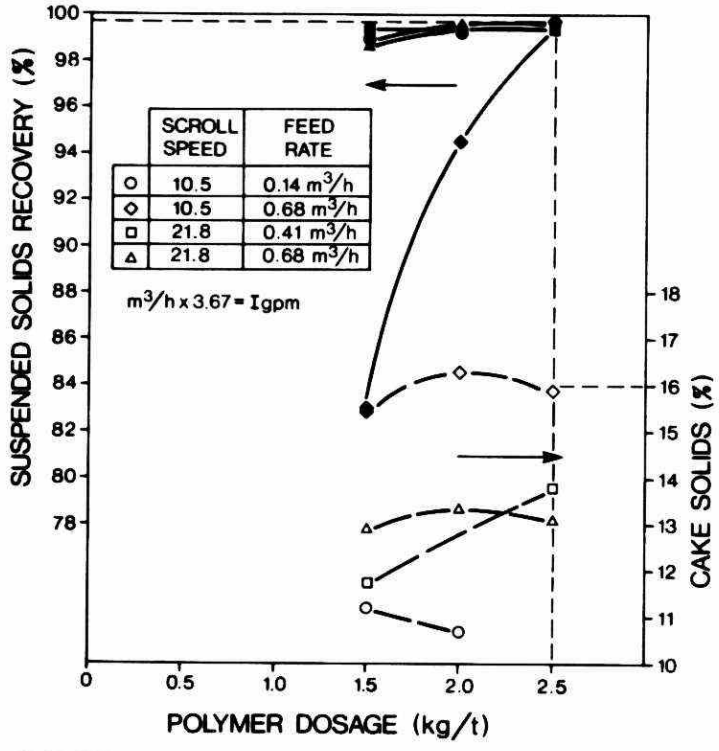


FIGURE 5-13. PERFORMANCE OF PILOT PLANT SOLID BOWL CENTRIFUGE

For the pilot scale centrifuge, the pool volume (V), can be calculated according to the following equation:

$$V = \pi \ell (r_2^2 - r_1^2) \qquad (5\text{-}15)$$

where:

$r_2$ = radius of bowl = 7.6 cm
$r_1$ = radius to inner surface of liquid = 6.0 cm
$\ell$ = bowl length = 30.5 cm

Substituting into:

$$V = \pi \ell (r_2^2 - r_1^2) \qquad (5\text{-}15)$$

$$= \pi(30.5)(7.6^2 - 6^2) = 2085 \text{ cm}^3$$

Additional variables required for the calculation of sigma are as follows:

$$\omega = \frac{4000 \times 2\pi}{60} = 420 \text{ radians/s}$$

$$g = 980 \text{ cm/s}^2$$

Substituting into:

$$\Sigma_p = \frac{V\omega^2}{g \ \ell n\left(\frac{r_2}{r_1}\right)} \qquad (5\text{-}11)$$

$$= \frac{2085 \ (420)^2}{980 \ \ell n\left(\frac{7.6}{6.0}\right)} = 0.16 \times 10^7 \text{ cm}^2$$

The remaining scale-up variables are as follows:

$$Q_{HP} = 0.68 \text{ m}^3/\text{h}$$
$$Q_{HF} = 25.0 \text{ m}^3/\text{h}$$

Substituting into:

$$\frac{Q_{HP}}{\Sigma_P} = \frac{Q_{HF}}{\Sigma_F} \qquad (5\text{-}13)$$

$$\Sigma_F = \frac{25.0 \times 0.16 \times 10^7}{0.68} = 5.9 \times 10^7 \text{ cm}^2$$

The full scale machine selected must have a theoretical hydraulic capacity factor of at least $5.9 \times 10^7$ cm$^2$ in order to achieve the specified performance criteria. Selection of the actual centrifuge

model would normally be made in consultation with the manufacturer. The sigma value determined above also represents a rational design parameter on which to base preliminary cost estimates.

DESIGN SUMMARY

| | |
|---|---|
| Flow Rate | 25 $m^3/h$ (90 Igpm). |
| Feed Solids Concentration | 6%. |
| Sludge Quantity | 13 140 tonne/year. |
| Scroll Differential Speed | 10.5 rpm. |
| Centrifugal Force | 1360 x G. |
| Hydraulic Capacity Factor ($\Sigma$) | 5.9 x $10^7$ $cm^2$. |
| Polymer Addition | 2.5 kg/t (5.0 lbs/ton). |
| Cake Solids | 16%. |
| Solids Recovery | 99%. |

(b) *Imperforate Bowl Basket Centrifuge*

SLUDGE CHARACTERISTICS

1. Waste activated sludge.
2. Flow rate = 25 $m^3/h$ (90 Igpm).
3. Feed solids concentration = 0.9%.

PERFORMANCE CRITERIA

1. Cake solids = 10%.
2. Solids recovery $\geq$ 99%.

A series of tests on a 35.6 x 15.2 cm (14 x 6 in) basket centrifuge produced the results shown in Figure 5-14. Bowl speed (2750 rpm, 1500 x G) was based on the manufacturer's recommendations and was not a test variable. The performance criteria could be satisfied (Figure 5-14) under the following conditions:

Flow rate = 0.41 $m^3/h$ (1.5 Igpm).
Polymer dosage = 1.25 kg/t (2.5 lbs/ton).

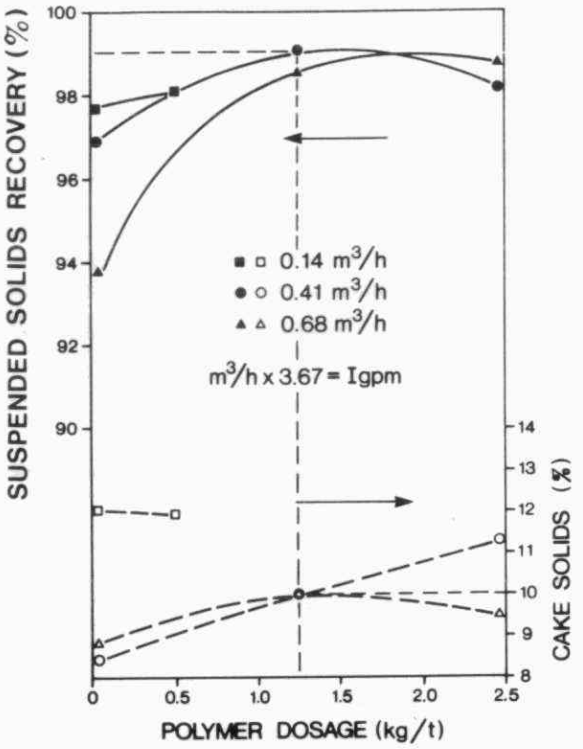


FIGURE 5-14. PERFORMANCE OF PILOT SCALE BASKET CENTRIFUGE

The usable volume of the pilot centrifuge basket was approximately 0.0085 $m^3$. A full scale model, 122 x 76 cm (48 x 30 in), recommended by the manufacturer for wastewater sludges, has a usable volume of approximately 0.34 $m^3$. The scale-up factor based on bowl volume would be:

$$0.34/0.0085 = 40$$

The density of the cake ($\rho_c$) can be calculated using the following equation:

$$\rho_c = \frac{\rho_d}{\rho_d - w(\rho_d - 1)} \qquad (5\text{-}16)$$

where: $\rho_d$ = dry solids density of the feed sludge = 1.59 g/cm$^3$

$w$ = fractional dry solids content of the cake = 0.1 (i.e., 10%).

Substituting into:

$$\rho_c = \frac{\rho_d}{\rho_d - w\,(\rho_d - 1)} \qquad (5\text{-}16)$$

$$= \frac{1.59}{1.59 - 0.1\,(1.59 - 1.0)} = 1.04 \text{ g/cm}^3 = 1040 \text{ kg/m}^3$$

For a full scale model, the weight of wet cake which will be processed per cycle is:

$$0.34 \text{ m}^3/\text{cycle} \times 1040 \text{ kg/m}^3 = 354 \text{ kg/cycle}$$

The weight of dry cake per cycle (cake solids = 10%) is:

$$354 \text{ kg/cycle} \times 0.1 = 35.4 \text{ kg/cycle}$$

The weight of wet feed sludge per cycle (feed solids = 0.9%) is:

$$\frac{35.4 \text{ kg/cycle}}{0.009} = 3933 \text{ kg/cycle}$$

Assuming the density of the wet feed sludge is 1.0 g/cm$^3$ (1000 kg/m$^3$), the volume of wet sludge processed per cycle is:

$$\frac{3933 \text{ kg/cycle}}{1000 \text{ kg/m}^3} = 3.9 \text{ m}^3/\text{cycle}$$

Flow rate can be scaled up on the basis of bowl volume ratios such that:

$$\text{Full scale flow rate} = 40 \times 0.41 \text{ m}^3/\text{h}$$
$$= 16.4 \text{ m}^3/\text{h}$$

Assuming the downtime at the end of each cycle is 2.5 minutes (varies with the manufacturer), the total operating cycle time is:

$$\left[\left(\frac{3.9\ m^3}{cycle}\right) \div \left(\frac{16.4\ m^3}{h} \times \frac{h}{60\ min}\right)\right] + \left[\frac{2.5\ min}{cycle}\right] = 16.8\ min/cycle$$

The average capacity of the machine (in terms of wet feed sludge) is:

$$\left[\left(\frac{60\ min}{h}\right) \div \left(\frac{16.8\ min}{cycle}\right)\right] \times \left(\frac{3.9\ m^3}{cycle}\right) = 13.9\ m^3/h$$

The number of machines required is:

$$\frac{25\ m^3/h}{13.9\ m^3/h} = 1.8$$

The installation of two machines would provide a measure of reserve capacity.

DESIGN SUMMARY

| | |
|---|---|
| Flow Rate | 25 $m^3/h$ (90 Igpm). |
| Feed Solids Concentration | 0.9%. |
| Sludge Quantity | 1971 tonne/year. |
| Number of Centrifuges | 2. |
| Size of Bowl | 122 x 76 cm (48 x 30 in). |
| Centrifugal Force | 1500 x G. |
| Polymer Dosage | 1.25 kg/t (2.5 lbs/ton). |
| Maximum Flow Rate/Unit | 16.4 $m^3/h$ (60 Igpm). |
| Cake Solids | 10%. |
| Solids Recovery | 99%. |

## 5.4 Vacuum Filtration

### 5.4.1 Process description and theory

Vacuum filtration is one of the oldest and most popular methods of mechanically dewatering sewage sludge. The component parts of a vacuum

filter are the filter drum, the filter media, a tank to hold the slurry, a filtrate pump and a vacuum pump (Figure 5-15). The drum, which is covered with the filter media, is partially submerged in the slurry tank. A vacuum is drawn inside the drum and lifts the sludge onto the filter surface. The liquid is drawn through the media and the solids retained on the outside. This is known as the form portion of the filtration cycle. As the drum rotates a given section moves out of the sludge slurry and the dry cycle begins. A vacuum continues to draw water from the cake formed on the filter media. At the end of this portion of the cycle, the cake is discharged, the filter media washed and the cycle started again. The most common types of filter media include various synthetic cloths and stainless steel coils. Discharge of the cake is normally accomplished either by a stationary scraper or by passing the filter media over a set of small rollers.

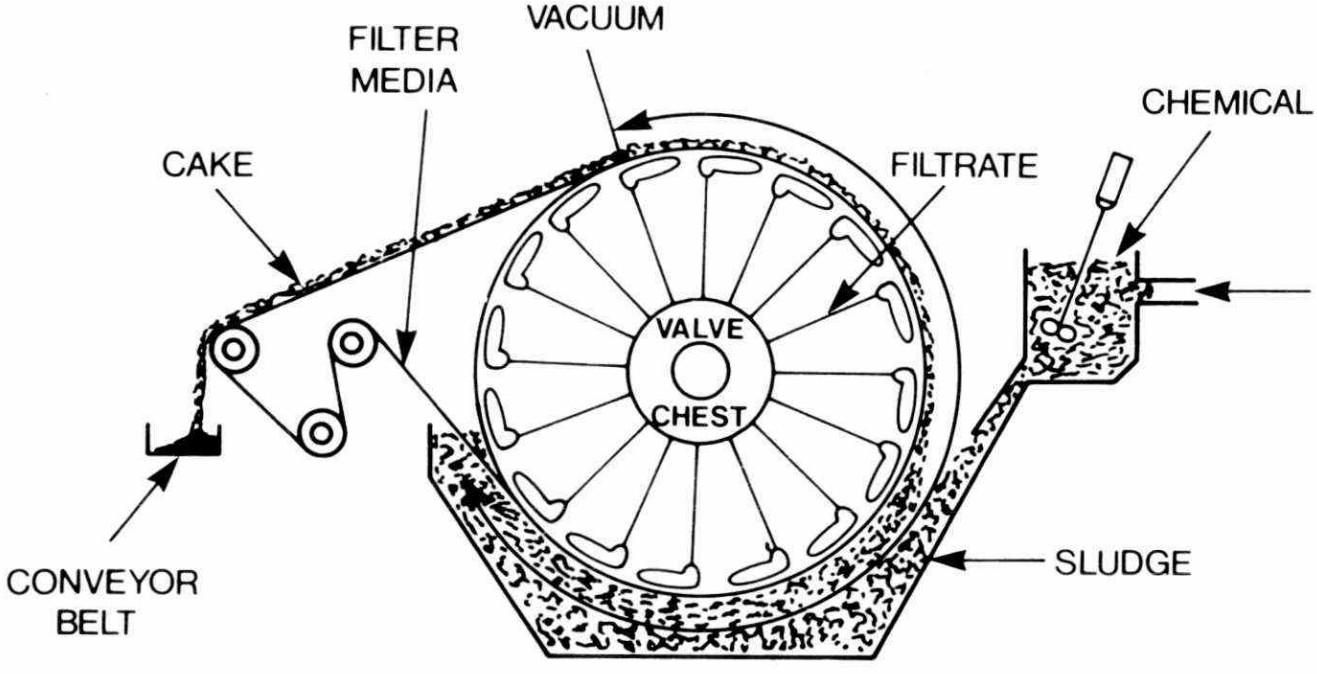


FIGURE 5-15. ROTARY VACUUM FILTER (REFERENCE: U.S. EPA, 1974)

The basic filtration equation is developed from the theory of liquid flow through a porous media. For the case of sewage sludges which form compressible cakes (assuming the resistance of the filter media is negligible) the equation is:

$$\frac{dv}{d\theta} = \frac{P}{\mu \bar{r}^* C^* v} \qquad (5\text{-}17)$$

where: $\frac{dv}{d\theta}$ = rate of filtrate flow per unit time ($m^3/m^2$-s).

$P$ = pressure drop ($N/m^2$).

$\mu$ = filtrate viscosity ($N\text{-}s/m^2$).

$\bar{r}^*$ = average value of the true specific resistance (m/kg).

$C^*$ = mass of dry suspended solids deposited on filter cake per unit volume of filtrate obtained ($kg/m^3$).

$v$ = volume of filtrate obtained from a unit area of filtration surface ($m^3/m^2$).

Integration and rearrangement of the equation permits the calculation of the specific resistance according to:

$$r = \frac{2 P A^2}{\mu C} b \qquad (2\text{-}2)$$

Equation 2-2 has been presented in Section 2.7 complete with a definition of the symbols and a discussion of specific resistance. A detailed explanation of filtration theory is available in the literature (Gale, 1967).

The performance of a vacuum filter, like all other dewatering processes, can be significantly affected by the choice of operating variables. The loading or throughput capacity to a filter is not an operating variable as it is in other processes. The capacity of a filter is termed the filter yield and is a function of the feed solids concentration, the filter drum speed, the ratio of form time to dry time and the degree of conditioning. Once the levels of these other variables have been established an increase in the feed rate to the filter will only result in the extra

sludge being discharged through the filter vat overflow pipe. The yield can be improved by increasing either the *feed solids concentration*, the *filter drum speed*, the *ratio of form time to dry time* or the *level of conditioning*. The cake solids concentration can be increased by decreasing either the *drum speed* or the *feed solids concentration*. The effect of conditioning on cake solids is not consistent and depends upon the type of conditioner used. The solids recovery is increased by either slower *drum speeds*, *low porosity cloths* or increased *conditioning*. Operational requirements of the vacuum filter system may limit the allowable changes in any variable to a very small range.

### 5.4.2 Development of design information

Vacuum filters can be designed either from bench test data or on the basis of pilot plant operation. Regardless of the method used to obtain the data, the design procedure involves determining the maximum filter yield that can be achieved under the optimum set of operating conditions. This value can in turn be related to the vacuum filter surface area required.

The bench tests commonly used for predicting filter operation are the capillary suction time test (CST), the Buchner funnel test, and the leaf filter test. All three tests provide a measure of the filtration characteristics of a sludge.

The CST (Baskerville and Gale, 1968) provides a measure of the *relative dewaterability* of a sludge and is most applicable for the evaluation of conditioning chemicals and methods. A good correlation between CST and specific resistance can be obtained for most sludges. The CST is used primarily for process monitoring as opposed to design.

The Buchner funnel test can be used to determine the *specific resistance*, the *coefficient of compressibility*, and for the prediction of *filter yield* (Eckenfelder and Ford, 1970). It has been extensively used as a method of determining sludge conditioning requirements despite the fact that it can be a very time consuming test and requires considerable data manipulation. Filter yields based on Buchner funnel data have been shown to give predictions within ±30% of the measured yields (Gale and Baskerville, 1970). Several equations for determining filter yield based on specific resistance values are available in the literature (Gale, 1971a; Adams and Eckenfelder,

1974; Vesilind, 1974c). Design based on specific resistance values is not recommended.

A potential problem associated with using Buchner funnel data for vacuum filter design purposes is that the test cannot evaluate the pick-up characteristics of the sludge. The possibility of a sludge having a relatively low specific resistance without the necessary physical characteristics to ensure formation of a cake on the filter is greater with polymers than with inorganic chemicals.

The leaf filter test (Eckenfelder and Ford, 1970) provides the best approximation of actual filter operation of the three tests described. The leaf filter test can be used to evaluate the effect of *conditioning, cycle time, ratio of form to dry time, pressure,* and *filter media* on the yield of the filter. The filter yield can be calculated according to the following equation:

$$\text{filter yield} = \frac{\text{weight of dry cake x number of cycles/hour}}{\text{filter area of the leaf}} \ (\text{kg/m}^2\text{-h}) \qquad (5\text{-}18)$$

To determine the optimum set of operating conditions it is possible to use a vacuum filter performance factor (PF) where:

$$PF = \frac{C_c}{\left(\frac{w}{w_s}\right)(Y)\ (R)} \qquad (5\text{-}19)$$

where: $C_c$ = conditioning cost ($/tonne of dry sludge).
$w$ = cake solids (%).
$w_s$ = feed solids (%).
$Y$ = filter yield (kg/m$^2$-h).
$R$ = solids recovery (% expressed as a fraction i.e., 95% = 0.95).

The optimum set of conditions is that which results in the lowest PF value (Cohen and Fullerton, 1975).

When leaf tests are conducted on polymer conditioned sludge it is important to consider the mechanical method of mixing the polymer in the

sludge. Recent work (Campbell et al, 1975) has shown that the yields obtained with sludge from the pilot plant conditioning tank were up to 2.5 times greater than those obtained on sludge which had the same polymer dosage, but had been mixed by hand in the lab. Although the methodology of flocculator scale-up is not well developed, an effort should be made to simulate full scale mixing conditions to the greatest possible extent when conducting bench scale experiments.

The operation of a pilot plant to generate design data for full scale equipment is probably less of a requirement for vacuum filtration than most other dewatering processes. This is due to the degree of confidence that can be placed on data obtained from leaf filter tests and the considerable amount of data that is available on the design and operation of full scale vacuum filter installations. However, pilot plant operation can provide insight into aspects of operation that are difficult to assess at the bench scale level. For example, sludge pickup and discharge characteristics of various types of filter media are best evaluated at pilot plant scale. Continuous operation will also provide information on the quantity and quality of the filter wash water and establish maximum allowable drum speeds.

### 5.4.3 Design example

SLUDGE CHARACTERISTICS

1. Digested sludge (elutriated).
2. Flow rate = 25 $m^3/h$ (90 Igpm).
3. Feed solids concentration = 5.7% (57 $kg/m^3$).
4. Dry solids flow = 25 $m^3/h$ x 57 $kg/m^3$
   = 1425 kg/h (3135 lbs/h).

PERFORMANCE CRITERIA

1. Solids recovery $\geq$ 95%.
2. Minimize performance factor (PF).

A summary of the leaf filter test results is shown in Table 5-3. The dosage of ferric chloride (6%) was based on the dewatering experience of the plant where the sludge sample was obtained. In

TABLE 5-3. LEAF FILTER TEST RESULTS[1]

| Filter Cloth | $FeCl_3$ % | Lime % | Cycle Time min/cycle | Cake Solids % | Filter[2] Yield $kg/m^2$-h | Chemical[3] Cost $/metric ton | Vacuum Filter Performance Factor PF |
|---|---|---|---|---|---|---|---|
| Nylon, | 6 | 6 | 5 | 17.3 | 16.8 | 14.04 | 0.275 |
| Low Porosity | 6 | 9 | 5 | 17.6 | 30.0 | 15.36 | 0.166* |
| " | 6 | 9 | 3 | 16.7 | 31.1 | 15.36 | 0.169 |
| " | 6 | 9 | 1.5 | 16.3 | 19.9 | 15.36 | 0.270 |
| " | 6 | 12 | 5 | 16.9 | 17.6 | 16.68 | 0.320 |
| " | 6 | 12 | 3 | 16.5 | 29.3 | 16.68 | 0.197 |
| " | 6 | 12 | 1.5 | 15.2 | 13.3 | 16.68 | 0.470 |
| " | 6 | 18 | 5 | 17.8 | 17.1 | 19.32 | 0.362 |
| " | 6 | 18 | 3 | 17.1 | 19.8 | 19.32 | 0.325 |
| " | 6 | 18 | 1.5 | 16.4 | 35.6 | 19.32 | 0.189* |
| Nylon, | 6 | 9 | 5 | 18.5 | 10.4 | 15.36 | 0.455 |
| High Porosity | 6 | 9 | 3 | 17.4 | 11.6 | 15.36 | 0.434 |
| " | 6 | 12 | 5 | 18.9 | 14.4 | 16.68 | 0.349 |
| " | 6 | 12 | 3 | 16.5 | 20.9 | 16.68 | 0.276 |
| " | 6 | 12 | 1.5 | 15.6 | 30.1 | 16.68 | 0.203 |
| " | 6 | 18 | 5 | 20.6 | 19.4 | 19.32 | 0.275 |
| " | 6 | 18 | 3 | 17.9 | 20.4 | 19.32 | 0.302 |
| " | 6 | 18 | 1.5 | 15.9 | 24.8 | 19.32 | 0.279 |
| Coil Leaf Filter | 6 | 12 | 5 | 17.5 | 16.4 | 16.68 | 0.331 |
| | 6 | 12 | 3 | 16.6 | 23.9 | 16.68 | 0.239 |
| " | 6 | 12 | 1.5 | 15.2 | 22.1 | 16.68 | 0.283 |

[1]Feed Solids = 5.7%.
[2]$kg/m^2$ x 0.205 = $lbs/ft^2$-h.
[3]December, 1975.
*Refer to text.

cases where polymers are an effective means of conditioning, polymer selection and the range of suitable dosages would normally be based on CST results. All tests achieved a solids recovery level of 95% to 97% and because of the limited range, this variable was not included in the calculation of the performance factor. Examination of Table 5-3 indicates the lowest performance factor was 0.166. This was obtained under the following conditions:

Conditioning = 6% $FeCl_3$, 9% lime.
Cycle time = 5 minutes.
Filter media = nylon, low porosity.

The corresponding cake solids and filter yield were 17.6% and 30.0 $kg/m^2$-h, respectively.

The required surface area is:

$$\frac{1425 \text{ kg/h}}{30.0 \text{ kg/m}^2\text{-h}} = 47.5 \text{ m}^2 \ (511 \text{ ft}^2)$$

A safety factor of 0.8 to account for changing sludge characteristics is commonly used to scale up from bench test data to full scale operation.

The required surface area is:

$$\frac{47.5}{0.8} = 59.4 \text{ m}^2 \ (640 \text{ ft}^2)$$

The development of a summary of results also allows the designer to evaluate the flexibility of his design. Using the filter designed above, it is possible to increase the capacity approximately 20%, by doubling the lime dosage from 9% to 18% and decreasing the cycle time from 5 to 1.5 min/cycle (Table 5-3). The performance factor of 0.189 under these conditions is not the optimum but is still the third lowest of all test conditions.

DESIGN SUMMARY

| | |
|---|---|
| Feed Rate | 25 $m^3/h$ (90 Igpm). |
| Feed Solids Concentration | 5.7%. |
| Solids Flow Rate | 1425 kg/h (3135 lbs/h). |
| Sludge Quantity | 12 483 tonne/year. |
| Filter Area | 59.4 $m^2$ (640 $ft^2$). |
| Cycle Time | 5 min/cycle. |
| Filter Media | Nylon, low porosity. |
| Conditioning | 6% $FeCl_3$, 9% lime. |
| Filter Yield | 30 $kg/m^2$-h (6.15 $lbs/ft^2$-h). |
| Cake Solids | 17.6%. |
| Solids Recovery | 95% to 97%. |

## 5.5 Pressure Filtration

### 5.5.1 Process description and theory

Pressure filtration has been widely used for many years in the chemical process industry for solid-liquid separation. It has been used extensively in Europe for dewatering wastewater sludges and is gaining acceptance in North America in applications where a very concentrated sludge cake (>30%) is required. The basic process is similar to vacuum filtration except that a positive pressure rather than a vacuum forces the liquid through the filter cloth. A filter press (Figure 5-16) consists of a number of plates and frames, or recessed plates, resting on some type of support. Sludge is pumped into the chambers formed by either the frames or the recesses in the plates. The plates are covered with a filter cloth such that as the chamber fills with sludge the liquid is forced through the cloth, travels through a drainage grid parallel to the cloth and to an outlet port. As sludge begins to compact within the chamber, the filtration rate decreases and at some predetermined rate, filtration is terminated. The plates are then separated, the sludge cake falls out and the cycle is ready to be initiated again.

The theoretical aspects of pressure filtration are based on the concepts of flow through porous media in much the same way as those of vacuum filtration. The pressing time of a full scale filter can be predicted

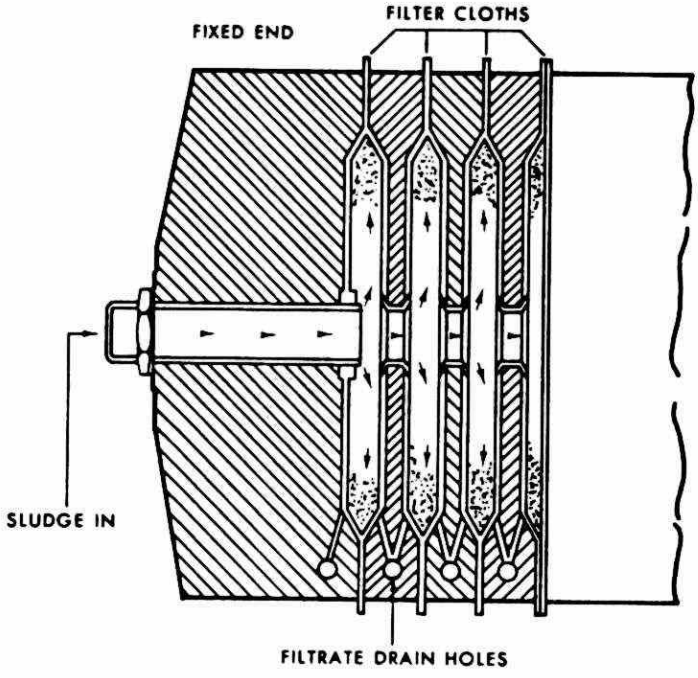


FIGURE 5-16. CUTAWAY VIEW OF A FILTER PRESS
(REFERENCE : U.S. EPA, 1974)

on the basis of specific resistance data obtained from the Buchner funnel test according to the following equation (Jones, 1956):

$$T = \frac{0.321\ \mu\ d^2\ r\ (C_i - C_f)^2}{P\ C_i\ (100 - C_i)} \qquad (5\text{-}20)$$

where:
- $T$ = pressing time (h).
- $\mu$ = filtrate viscosity (centipoise).
- $d$ = chamber thickness (in).
- $r$ = specific resistance x $10^7$ cm/g measured at 100 psi.
- $C_i$ = moisture content of the sludge (%).
- $C_f$ = moisture content of the cake (%).
- $P$ = pressure (psi).
- 0.321 = constant arising from the units system used.

Although the effect of variables on the pressing time as indicated by equation 5-20 is correct, the actual pressing times with a pilot scale unit were found to be on the average 2.35 times longer than those predicted (Baskerville et al, 1971).

The variables which most affect filter press performance are the *operating pressure*, the *chamber thickness*, and the *characteristics of the sludge*. The optimum operating pressure is related to the compressibility of the sludge and, therefore, for municipal sludges lower pressures ($\leq$ 690 kN/m$^2$) are normally used. The maximum chamber thickness, within equipment limits, which will allow a reasonable pressing time is usually selected. The characteristics of the sludge can be improved by conditioning such that the specific resistance is decreased. The use of inorganic conditioners such as fly ash and the practice of precoating with diatomaceous earth are common.

The advantage of pressure filtration lies in the combination of high cake solids with very low levels of suspended solids in the filtrate. The traditional disadvantages of the system have been related to the fact that it is a batch process and the shifting of the plates was a very labour intensive process. This has been largely overcome by the introduction of a mechanical plate shifter and a general increase in the automation of the system.

### 5.5.2 Development of design information

The prediction of filtrate quality is one aspect of design which is not normally a problem with pressure filtration. Due to the nature of the process, some solids are present in the initial stages of filtration but decrease rapidly as the chamber becomes full. The two objectives in developing data for design are to predict the cake density and length of the pressing cycle. As mentioned in the previous section Buchner funnel data could be used to obtain a prediction of the pressing time but this method is not recommended. The use of the Buchner funnel test should be limited to the evaluation of different methods of conditioning.

Bench scale filter presses provide information on cake density and pressing time but are generally not sufficiently accurate to be used

for design purposes. Most bench presses have only one filtering surface and are filled with sludge before filtration is started (Ackers and Moss, 1973). The parabolic shape of the resulting filtration curve (straight line on log-log plot) is indicative of unrestricted filtration. In a pilot plant the press must first be filled, during which time there is a continual increase in the filtration rate. Once the press is full, filtration occurs in a similar manner to the bench unit until buildup of the two cakes progresses to the point at which they interfere with each other. As this occurs a drop off in the filtration rate is evident. Figure 5-17 illustrates the difference between the filtration curves for unrestricted filtration (bench test) and two-sided filtration (pilot plant). Pressing times estimated on the basis of bench tests would tend to result in considerably lower cake solids than expected.

The bench test is an excellent method of predicting the range of cake solids and evaluating the effects of conditioners, filter media and operating pressure. Figure 5-18 illustrates how the test may be used to select the optimum polymer dosage. A polymer dosage of 2.125 kg/t resulted in the largest volume of filtrate in the shortest time. As the dosage was increased to 2.25 kg/t, the filtration rate decreased.

The most accepted method of generating design information is to operate a pilot plant. These pilot plants usually have several chambers which range from 30 to 60 cm square (12 to 25 inches). Data developed in this manner will provide a good prediction of cake solids, filtration rates and pressing times. The results can then be directly scaled-up to full scale.

### 5.5.3 Design example

SLUDGE CHARACTERISTICS

1. High lime sludge (base metal mine treatment waste).
2. Flow rate = 25 $m^3/h$ (90 Igpm).
3. Feed solids concentration = 7%.
4. Bulk density = 1.044 $g/cm^3$.
5. Dry solids density = 3.256 $g/cm^3$.

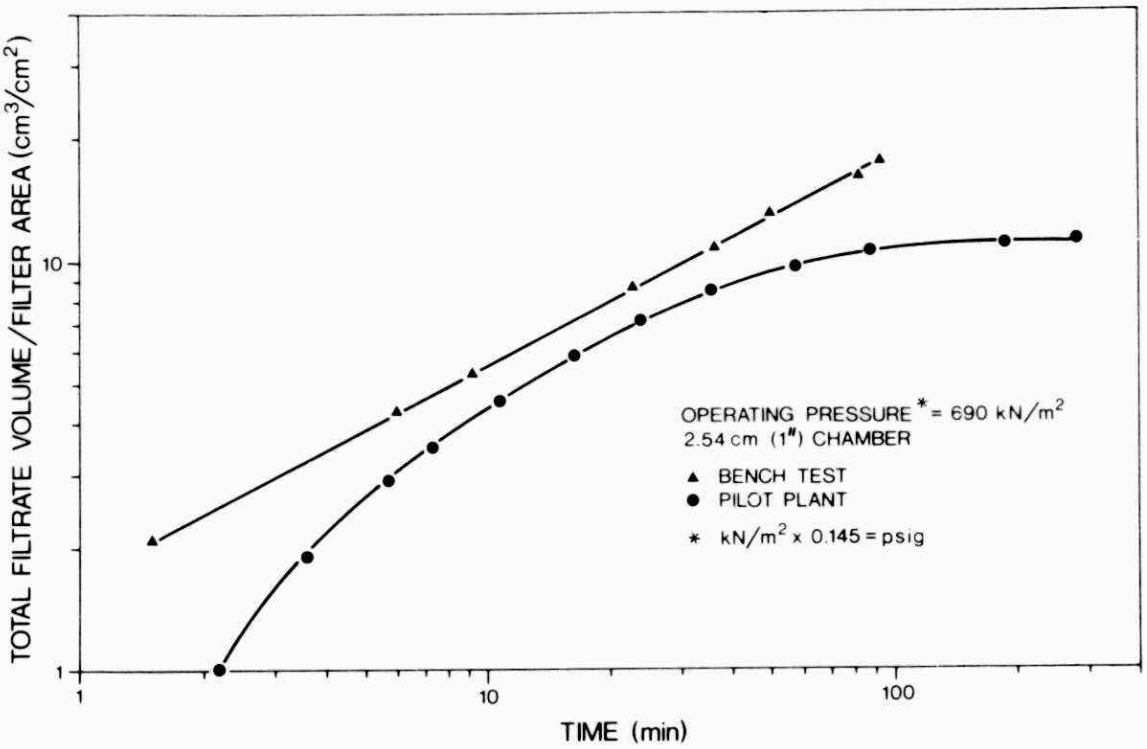


FIGURE 5-17. A COMPARISON OF BENCH AND PILOT SCALE PRESSURE FILTRATION OPERATION

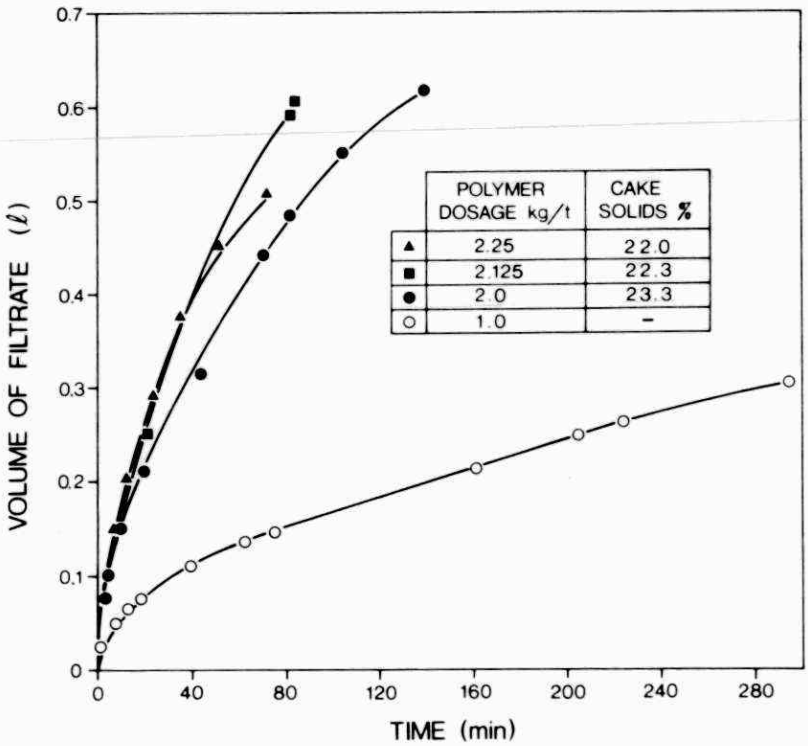


FIGURE 5-18. BENCH FILTER PRESS - COMPARISON OF CONDITIONING REQUIREMENTS

PERFORMANCE CRITERIA

1. Cake solids = 38%.
2. Solids recovery $\geq$ 99%.

Bench tests carried out at four different pressures (Figure 5-19) indicated that the effect of pressure was marginal. The highest flow rate was obtained at 690 kN/m$^2$ while the highest cake solids were obtained at 1035 kN/m$^2$.

A 45.7 x 45.7 cm (18 x 18 in) plate and frame pressure filter was operated using four chambers, to produce the results shown in Figure 5-20. A cotton filter media was used, but information on the porosity of the cloth was not available. Chemical conditioning was not required as all tests produced a filtrate with less than 10 mg/$\ell$ of suspended solids.

In Figure 5-20, the filtration rate ($\ell$/min) at any time is given by the slope of the curves shown. The highest filtration rates were achieved at a pressure of 690 kN/m$^2$ for the 2.54 cm frames and at 1035 kN/m$^2$ for the 5.08 cm frames. These two tests have been replotted separately in Figure 5-21. The end conditions of the test at 690 kN/m$^2$ were as follows (Figure 5-21):

Press time = 464 min.
Cake solids = 42.4%.
Total filtrate volume = 163 $\ell$.

Assuming the filtrate solids are negligible, the cake solids at any time ($t_1$) can be related to the cake solids at any other time ($t_2$) by considering a mass balance according to the following relationship:

$$\left[\frac{C}{V_s}\right]_{t_1} = \left[\frac{C}{V_s}\right]_{t_2} \qquad (5\text{-}21)$$

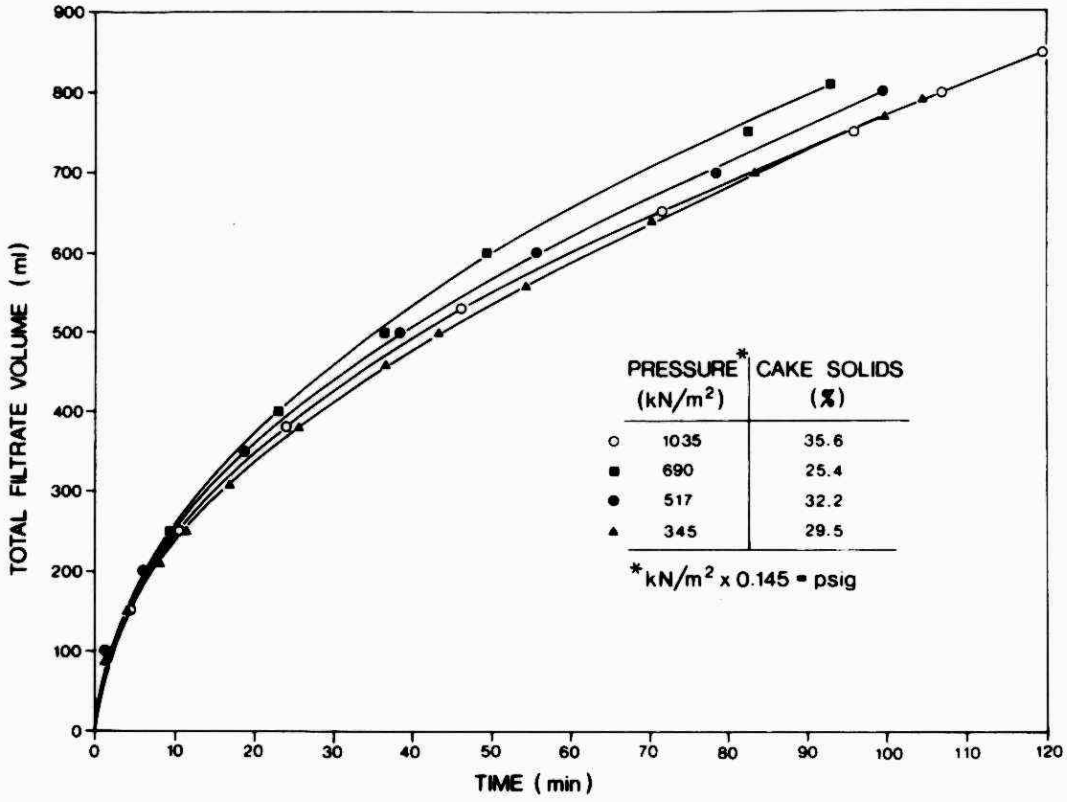


FIGURE 5-19. EFFECT OF OPERATING PRESSURE ON BENCH FILTER PRESS TESTS

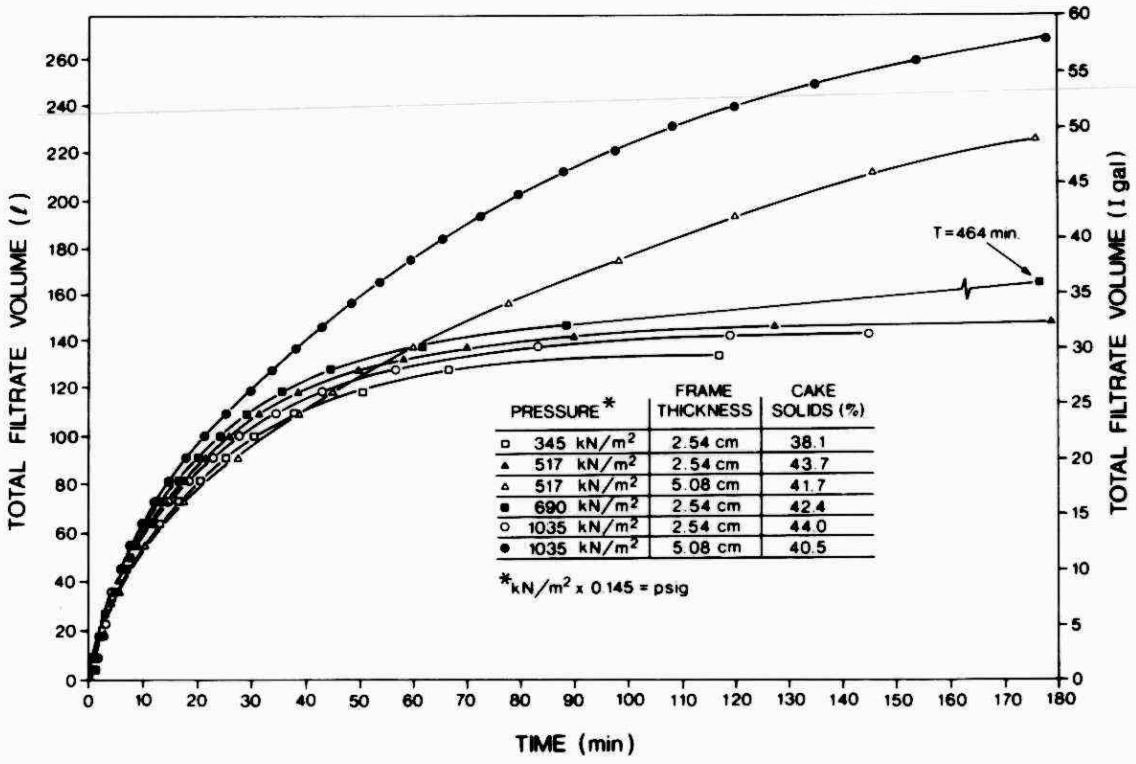


FIGURE 5-20. EFFECT OF OPERATING PRESSURE AND FRAME SIZE ON FILTRATE FLOW FROM FILTER PRESS

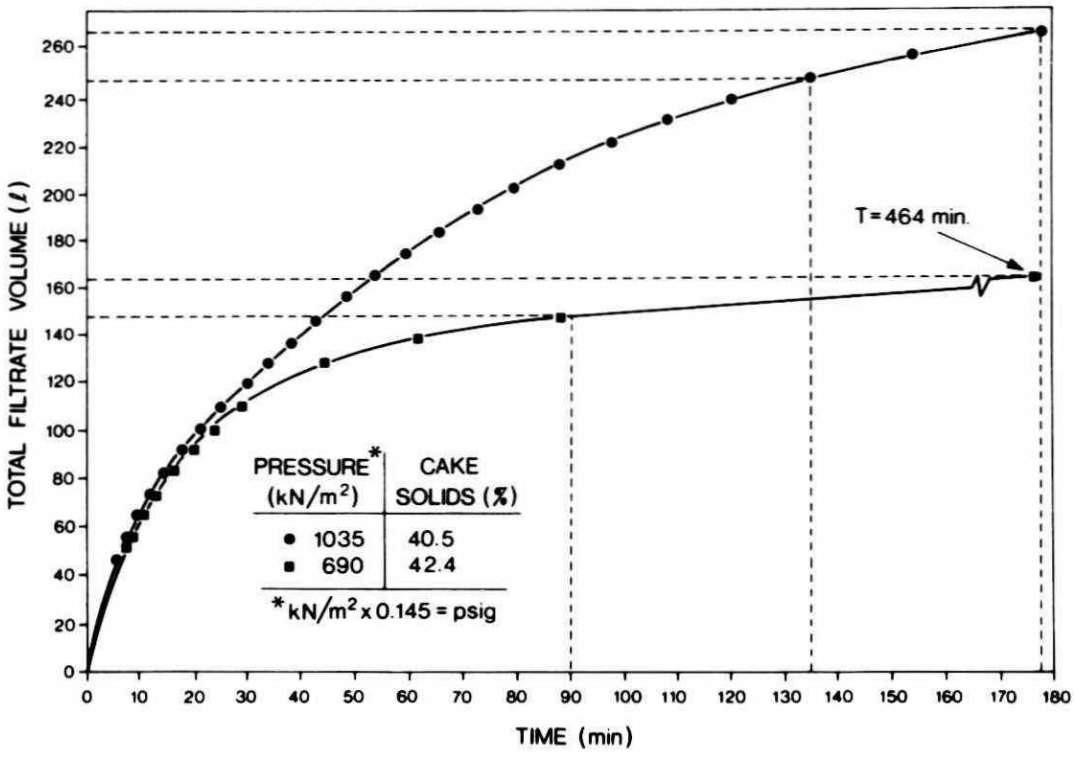


FIGURE 5-21. EFFECT OF OPERATING PRESSURE AND FRAME SIZE ON FILTRATE FLOW FROM FILTER PRESS

where: $C$ = cake solids (%).

$V_s$ = volume of sludge pressed (ℓ).

The volume of sludge pressed is determined by:

$$V_s = V_f + V_c \qquad (5\text{-}22)$$

where: $V_f$ = volume of filtrate (ℓ).

$V_c$ = volume of chamber (ℓ).

The chamber volume ($V_c$) for the pilot plant using 2.54 cm frames (chambers) was 17.9 ℓ.

To obtain a cake solids of 38%:

$C_2 = 42.4\%$.

$V_{s_2} = 163 + 17.9 = 180.9\ \ell$. (from Equation 5-22)

$C_1 = 38\%$.

Substituting into:

$$\left[\frac{C}{V_s}\right]_{t_1} = \left[\frac{C}{V_s}\right]_{t_2} \qquad (5\text{-}21)$$

$$V_{s_1} = \frac{38 \times 180.9}{42.4} = 162.1\ \ell.$$

According to:

$$V_s = V_f + V_c \qquad (5\text{-}22)$$

$$V_f = 162.1 - 17.9 = 144.2\ \ell.$$

Reference to Figure 5-21 indicates that at 690 kN/m$^2$ and 2.54 cm frames, 144.2 $\ell$ of filtrate were collected in 90 min.

Based on the assumption that 30 min of downtime are required at the end of each cycle:

Total cycle time = 90 + 30 = 120 min (2 h).

$$\text{Number of cycles/day} = \frac{24\ \text{h/d}}{2\ \text{h/cycle}} = 12\ \text{cycles/day}.$$

The density of the cake can be calculated according to equation 5-16:

$$\rho_c = \frac{\rho_d}{\rho_d - w\ (\rho_d - 1)} \qquad (5\text{-}16)$$

where: $\rho_c$ = density of the cake (g/cm$^3$).

$\rho_d$ = dry solids density of the sludge (3.256 g/cm$^3$).

w = fractional dry solids content of the cake (38% = 0.38).

Substituting into Equation 5-16:

$$\rho_c = \frac{3.256}{3.256 - 0.38\ (3.256 - 1)} = 1.357\ g/cm^3 = 1357\ kg/m^3$$

$$\text{Dry solids flow rate} = \frac{25\ m^3}{h} \times \frac{24\ h}{d} \times \frac{1044\ kg}{m^3} \times \frac{0.07\ kg}{kg}$$

$$= 43848\ kg\ dry\ solids/d$$

The volume of dewatered cake is:

$$\frac{43848\ kg/d}{1357\ kg/m^3} = 32.3\ m^3/d$$

The required press volume is:

$$\frac{32.3\ m^3/d}{12\ cycles/d} = 2.7\ m^3/cycle$$

Any combination of chamber size (area) and number of chambers which will provide a total volume of 2.7 m$^3$ (95.3 ft$^3$) in 2.54 cm chambers would be acceptable.

The end conditions of the test at 1035 kN/m$^2$ were as follows (Figure 5-21):

Press time = 178 min.

Cake solids = 40.5%.

Total filtrate volume = 264 ℓ.

The chamber volume ($V_c$) for the pilot plant using 5.08 cm frames was 35.7 ℓ.

To obtain a cake solids of 38%:

$C_2 = 40.5\%$.

$V_{s_2} = 264 + 35.7 = 299.7$ ℓ. (from Equation 5-22)

$C_1 = 38.0\%$.

Substituting into:

$$\left[\frac{C}{V_s}\right]_{t_1} = \left[\frac{C}{V_s}\right]_{t_2} \qquad (5\text{-}21)$$

$$V_{s_1} = \frac{38 \times 299.7}{40.5} = 281.2 \text{ ℓ}.$$

According to:

$$V_s = V_f + V_c \qquad (5\text{-}22)$$

$$V_f = 281.2 - 35.7 = 245.5 \text{ ℓ}.$$

Reference to Figure 5-21 indicates that at 1035 kN/m$^2$ and 5.08 cm frames, 245.5 ℓ of filtrate were collected in 135 min.

Based on the assumption that 30 min of downtime are required at the end of each cycle:

Total cycle time = 135 + 30 = 165 min (2.75 h).

$$\text{Number of cycles/day} = \frac{24 \text{ h/d}}{2.75 \text{ h/cycle}} = 8.7 \text{ cycles/day}.$$

The required press volume is:

$$\frac{32.3 \text{ m}^3\text{/d}}{8.7 \text{ cycles/d}} = 3.7 \text{ m}^3\text{/cycle}$$

Based only on a consideration of the size of equipment required, a press with 2.54 cm frames (2.7 m$^3$) would be selected in preference to one with 5.08 cm frames (3.7 m$^3$).

DESIGN SUMMARY

| | |
|---|---|
| Flow Rate | 25 $m^3/h$ (90 Igpm). |
| Feed Solids Concentration | 7%. |
| Solids Flow Rate | 43 848 kg/d (96 464 lbs/d). |
| Sludge Quantity | 15.987 tonne/year. |
| Filter Press Volume | 2.7 $m^3$ (95.3 $ft^3$). |
| Operating Pressure | 690 $kN/m^2$ (100 psig). |
| Chamber Thickness | 2.54 cm (1.0 in). |
| Pressing Time | 1.5 h. |
| Total Cycle Time | 2.0 h. |
| Chemical Conditioning | None. |
| Cake Solids | 38%. |
| Solids Recovery | 99%. |

## 5.6 Belt Filter Press

In the past few years a variety of new or modified sludge dewatering processes have become available which offer an alternative to conventional methods such as vacuum filtration and centrifugation. The device which has probably attracted the most attention is the belt filter press (Figure 5-22). A wide variety of mechanical configurations are available from the different manufacturers but the principle on which they operate is common to all units. The sludge is introduced between two filter belts which move over a series of rollers. The pressure exerted by the belts, on the sludge, becomes progressively greater as the belts move through the press. At the end of the press the two belts separate and the cake is discharged.

The performance of most units is affected by *sludge feed rate, belt speed, belt pressure, belt material* and *chemical conditioning*.

Very little information is reported in the literature on the methodology used to determine full scale requirements. Most manufacturers' brochures express the capacity of their belt filter presses in terms of either total volume of sludge throughput or weight of dry solids throughput.

Bench tests are not available for the generation of design data. At least one manufacturer has reported using a hand squeezing test. This

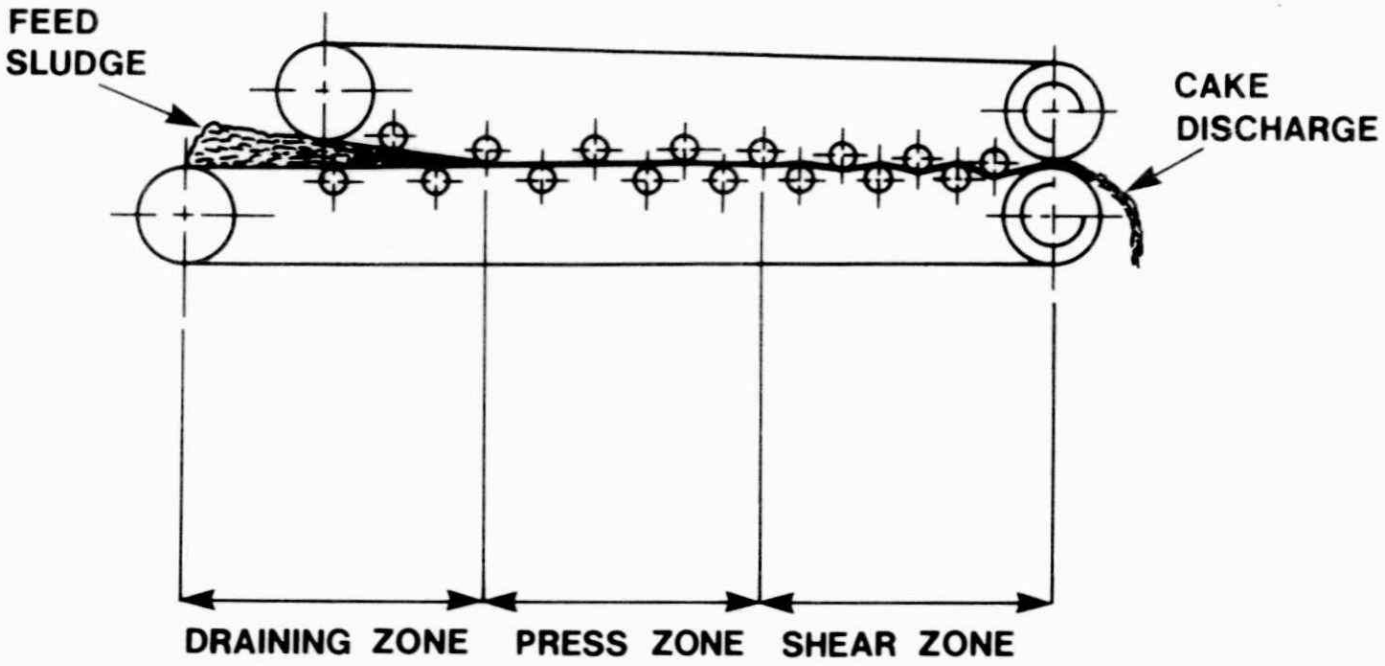


FIGURE 5-22. BELT FILTER PRESS (REFERENCE: U.S. EPA, 1974)

permits a qualitative evaluation of whether it will tend to extrude through the filter media under application of pressure by the belts.

The pilot scale machines used to test a sludge are usually the smallest full scale machine made by the manufacturer. Pilot scale and full scale units are identical with the exception of belt width which may range from 0.5 to 3.0 m. Experience has shown that scale-up is not linear with respect to belt width because as the width increases, the edge effect, or amount of unused belt area, becomes a progressively smaller proportion of the total surface area. At the present time the only way to determine full scale size requirements is to have a pilot plant machine tested on site. Based on the performance achieved by the small machine, recommendations as to expected full scale performance and total equipment requirements can be made.

# 6 COST ESTIMATION METHODOLOGY

The selection of a sludge dewatering system for any specific installation is usually based on two factors, the required performance level and the system cost. The required performance level is often dictated by circumstances outside the control of the designer. Bench and pilot plant tests can determine those systems which satisfy these requirements. Process selection is then reduced to estimating the cost of each system and choosing the one with the lowest cost. The time required to perform a cost estimate and the resulting accuracy can vary widely and should be a function of the intended use of the estimate.

A high degree of confidence is an essential feature of the cost estimates prepared by contractors prior to submitting contract bids. At this stage a detailed flow diagram is usually available such that all pieces of equipment, piping, fittings, etc., can be determined. Using delivered costs for each item, and unit installation costs the estimator can develop a detailed cost estimate. The method is extremely time consuming and as a result large companies employ experienced full time cost estimators for this job. This type of cost estimation is beyond the scope of this manual.

The most useful type of cost estimate for selecting a dewatering process is one which can be completed in an hour or two in order to obtain "ballpark" figures. This type of analysis is usually developed from published cost curves for the essential pieces of equipment and the associated operation and maintenance costs for the system. Woods (1975) suggests that analyses of this nature should be within ± 30% but, depending upon the source of the cost data, may be significantly more accurate than the ± 30%. When used in the following context, an estimate with this degree of error is quite adequate. Preliminary cost estimates provide a means of comparing two or more sludge dewatering alternatives on the basis of total cost; they give an estimate of the magnitude of capital investment required; and they provide a breakdown between the capital costs, and operation and maintenance costs, associated with each process.

One of the greatest problems in obtaining a quick estimate with reasonable accuracy, is relating costs to design parameters. Much of the cost data available for sludge dewatering is a function of the raw sewage flow to the treatment plant. If the sludge production or dewatering characteristics deviate from the average values assumed, this will not be reflected in the cost estimate. Where possible, cost data should be a function of a design parameter for the dewatering process in question. Even when this type of information is available caution must be exercised in both the development and application of the cost estimate. The costs presented in this manual are in terms of design parameters, and represent average costs, but are not necessarily representative of the lowest price available from other manufacturers for similar equipment.

## 6.1 Cost Indices

Cost indices are simply a way of expressing the effect of inflation on the cost of construction. The average costs of a specified set of factors including material prices and wages, are related to the cost of the same items at some arbitrary base year where the costs are usually assigned a value of 100. Actual cost data reported in the literature for any year, e.g., 1970, can then be adjusted using a cost index to predict the cost of the same facilities in any other year, e.g., 1976. Table 6-1 presents the values of three indices commonly used for updating construction costs.

The two most popular indices in the wastewater treatment field have traditionally been the Engineering News Record-Construction Cost Index (ENR-CCI) and the U.S. Environmental Protection Agency - Sewage Treatment Plant Index (EPA - STP). The Environmental Protection Agency has recently introduced two new indices, Small City Conventional Treatment (SCCT) and Large City Advanced Treatment (LCAT) to replace the older STP index (Winklehaus, 1976). These indices are expected to be more accurate than the old one due to a more detailed breakdown in the cost components. The EPA and Engineering News Record indices are suitable for use in total plant construction cost estimates because they contain a large component of structural building materials. For more specialized applications such as

TABLE 6-1. AVERAGE ANNUAL VALUES FOR COST INDICES

| Year | Engineering News-Record Construction Cost Index 1913 = 100 | | | Environmental Protection Agency | | | Marshall and Stevens Installed Equipment Index 1926 = 100 |
|---|---|---|---|---|---|---|---|
| | U.S. National Average | Toronto | Montreal | STP 1957-59 = 100 | SCCT 1973 3rd Quar. = 100 | LCAT 1973 3rd Quar. = 100 | |
| 1960 | 824 | 690 | 660 | 105 | - | - | 238 |
| 1961 | 847 | 701 | 678 | 106 | - | - | 237 |
| 1962 | 872 | 724 | 707 | 107 | - | - | 239 |
| 1963 | 901 | 743 | 736 | 109 | - | - | 239 |
| 1964 | 936 | 782 | 756 | 110 | - | - | 242 |
| 1965 | 971 | 844 | 785 | 112 | - | - | 245 |
| 1966 | 1021 | 887 | 930 | 116 | - | - | 252 |
| 1967 | 1070 | 922 | 1027 | 119 | - | - | 263 |
| 1968 | 1165 | 1050 | 1066 | 124 | - | - | 273 |
| 1969 | 1272 | 1082 | 1211 | 133 | - | - | 285 |
| 1970 | 1418 | 1183 | 1250 | 144 | - | - | 303 |
| 1971 | 1620 | 1334 | 1318 | 160 | - | - | 321 |
| 1972 | 1670 | 1536 | 1415 | 172 | - | - | 332 |
| 1973 | 1896 | 1763 | 1608 | 183 | 100 | 100 | 334 |
| 1974 | 2020 | 1913 | 1890 | 210 | - | - | 398 |
| 1975 | | | | | | | |
| 1st Quar. | 2163 | 2007 | 1952 | 247 | - | - | 435 |
| 2nd Quar. | 2248 | 2076 | 2086 | 246 | - | - | 442 |
| 3rd Quar. | 2293 | 2220 | 2093 | 251 | 109 | - | 447 |
| 4th Quar. | 2297 | 2240 | 2094 | 255 | 110 | 120 | 451 |
| 1976 | | | | | | | |
| 1st Quar. | 2327 | 2269 | 2141 | - | 113 | 123 | 460 |
| 2nd Quar. | 2413 | 2461 | 2273 | - | 115 | 126 | 469 |
| 3rd Quar. | 2478 | 2461 | 2273 | - | 117 | 129 | 477 |
| 4th Quar. | 2494 | 2530 | 2336 | - | 119 | 130 | 485 |

sludge dewatering systems, which are highly equipment oriented, cost updating should be based on a more appropriate index. It has been suggested (Woods, 1975) that the Marshall and Stevens Index (MS Index) may be more applicable due to the inclusion of a 25% factor for process equipment. Updated values for the cost indices are available in Engineering News Record (ENR-CCI), Journal Water Pollution Control Federation (EPA - STP, SCCT, LCAT) and Chemical Engineering (MS Index). A detailed discussion of the components included in the major indices is available in the literature (Woods, 1975).

## 6.2 Cost Components

The costs associated with a sludge dewatering installation fall into two categories, the fixed capital investment and the operation and maintenance costs. Methods of estimating each of these costs are presented in the following sections.

### 6.2.1 Capital investment

An estimate of the capital investment can be developed using several approaches. One of the most common approaches is to obtain values for the total capital investment from cost curves in the literature. This can be risky when comparing alternatives because many of the cost curves do not include the same components. Since these are total cost curves it is impossible to separate out an item such as a building which may not be required for the installation under consideration. Another method is to obtain FOB prices for the major pieces of equipment and develop a value for the total fixed capital investment based on factors which relate the installation costs to the cost of the equipment.

The factor method of cost estimation used in this manual is based on information presented by Woods (1975). The equipment costs were obtained from manufacturers and represent a total package which includes the dewatering unit and all major auxiliary equipment, but does not include feed pumps or sludge conveyors. Unit cost curves for each process and for sludge feed pumps are presented in the following sections. The cost of conveyors was not included since this cost is too site specific. The total FOB cost of the equipment is assigned a value of 1.0.

The next major expense is the installation of the equipment. Here the definition of what constitutes installation is critical. In this context it is the total labour and materials required for the placement and hookup of the dewatering equipment including piping, structural steel, foundations and reinforced concrete, insulation and painting, electrical, instrumentation, and site preparation. It does not include the cost of buildings or land and assumes that reasonable subsoil conditions prevail. Individual factors for installation of the components are available for more detailed cost analyses (Woods, 1975). For the purpose of preliminary cost estimates it is sufficient to group all of the installation materials and labour into a single factor. Installation factors for the various dewatering processes presented in Table 6-2 (Woods, 1976) range from 1.5 to 2.0. The complete installed cost (equipment cost plus installation cost) of a gravity thickener for example, is equal to 1.6 times the FOB cost of the equipment.

TABLE 6-2. INSTALLATION FACTORS[1] FOR DEWATERING EQUIPMENT

| Unit Process | Factor |
|---|---|
| Gravity Thickening | 1.6 |
| Dissolved Air Flotation | 1.5 |
| Solid Bowl Scroll Centrifuge | 1.6 |
| Basket Centrifuge | 1.57 |
| Vacuum Filter | 1.6 - 2.0 |
| Plate & Frame Pressure Filter | 1.79 |

[1]Does not include buildings.

Taxes, freight and insurance are assumed to be approximately 8% of the equipment cost. The sum of the equipment cost plus installation cost plus taxes, freight and insurance represents the total physical plant cost.

A bid submitted by a contractor to a municipality would include one other component, the contractor's fees. These represent the contractor's

profit and overhead, and may range from 10% profit, 10% overhead (based on the total physical plant cost) on small jobs, to a total of 5% for both, on very large jobs. The sum of the total physical plant cost plus the contractor's fees will be referred to as the contract cost throughout the remainder of the manual.

The second component of the fixed capital investment can be referred to as the indirect costs and includes engineering, field expenses, legal and other fees, contingencies, interest during construction, and working capital. The values suggested for these components are based on personal communication with contractors, consulting engineers, and Woods (1976), and are identified as follows:

Engineering services are usually provided by a consulting engineers' firm and include the engineering man-hours and overhead costs required to design the dewatering process. This expense may range from 7% to 12% of the contract cost depending upon the complexity of the project.

Field expenses refer to the costs to the contractor of on-site equipment such as field offices and includes construction supervision, purchasing, and liability insurance. This expense is assumed to be 5% of the contract cost.

Legal, fiscal and administration costs are related to the preparation of legal documents and construction contracts, and the arrangement of debt financing. This cost is assumed to be 2% of the contract cost.

Contingencies are an allowance for unforeseen events and are usually assumed to be 6% to 7% of the contract cost.

Interest during construction is the interest cost on funds required for interim payments to the contractor prior to the completion of construction and is assumed to be 6% of the contract cost.

Working capital refers to the cost of the initial order of supplies required for start-up of the plant. In the case of a dewatering installation this would include such items as polymer or other conditioning chemicals. This cost is small and is assumed to be 1% to 2% of the contract cost.

Start-up expense refers to the cost of testing and operation for an initial period of time to demonstrate that the process meets the

performance specifications. This cost is most commonly encountered with the installation of either an unproven process or an unproven application of a conventional process. The cost is very specific to the individual situation and as a result no generalized values are reported.

The sum of the contract cost and the indirect costs represents the total capital investment or bonded capital cost. An example using the installation factor for gravity thickening will serve to illustrate the method outlined above.

| | Cost Component | | Factor |
|---|---|---|---|
| 1. | Total equipment cost FOB | 1.0 | |
| 2. | Installation factor | 1.6 | |
| 3. | Installed cost | 1.6 x 1.0 | 1.6 |
| 4. | Tax, freight, and insurance | 0.08 x 1.0 | 0.08 |
| 5. | *Total Physical Plant Cost* | | 1.68 |
| 6. | Contractor's fees | 12% of (5) | 0.20 |
| 7. | *Contract Cost* | | 1.88 |
| 8. | Engineering expense | 10% of (7) | 0.19 |
| 9. | Field expense | 5% of (7) | 0.09 |
| 10. | Legal and other | 2% of (7) | 0.04 |
| 11. | Contingency | 6% of (7) | 0.11 |
| 12. | Interest during construction | 6% of (7) | 0.11 |
| 13. | Working capital | 1.5% of (7) | 0.03 |
| 14. | *Bonded Capital Cost* | | 2.45 |

Therefore, the total investment in this example, excluding buildings and land, is equal to 2.45 times the FOB cost of the major equipment.

The required capital investment funds for sewage treatment plant construction are often raised through the sale of municipal debentures. It is assumed that the interest rate paid on these debentures is equal to 10% and that the dewatering equipment has a 20-year life-span with no resale value. Under these circumstances the yearly cost of financing the construction is equal to 10.57% of the total fixed capital investment (calculated by the capital recovery depreciation method presented by Riggs, 1968).

### 6.2.2 Operation and maintenance costs

Operation and maintenance costs are subdivided into four sections: operating labour, maintenance labour, materials and supplies, and chemical costs. Operating labour includes the salaries of supervisory staff, operating personnel and general labourers. Maintenance labour represents the salaries of the maintenance staff. The total labour requirements are derived from a correlation between man-hours and amount of sludge processed for each type of equipment. Based on data presented in the 1975 Ontario Ministry of Environment operating summary, an average cost of $6.30 per hour (for Ontario) has been assumed for both operations and maintenance, to arrive at the total annual labour cost.

Materials and supplies include power and the normal materials required for maintenance of the equipment and plant. These are also based on the amount of sludge processed.

Chemical costs represent the cost of the chemicals used to condition the sludge where required. The use of generalized curves to estimate this cost is not recommended. The estimate should be based on the dosages determined in bench or pilot plant tests and the current cost of the specific chemicals used.

## 6.3 Unit Cost Curves

In this Section cost curves are presented for the equipment cost, and the operation and maintenance costs for the following unit operations: sludge pumping, gravity thickening, dissolved air flotation, solid-bowl centrifugation, basket centrifugation, vacuum filtration, and pressure filtration. Cost data presented for belt filter presses includes only equipment cost. The mathematical formula for each curve is available in the Appendix. Details on the equipment costs (December 1975) are explained for each curve. For the time base used to prepare the cost curves (December, 1975) the U.S. and Canadian dollar were essentially at par. Due to recent developments this is no longer true and must be considered when dealing with U.S. suppliers. Items which are not included in any curves are land costs, delivery costs, and conveyor costs. These costs are site specific and average values are meaningless. In certain situations they may represent a significant cost factor and should be evaluated for

each project. The explanation of the operation and maintenance costs was presented in Section 6.2.2. It must be noted that all labour requirements for operation and maintenance are presented in terms of man-hours per year and *must not* be adjusted by a cost index. The hourly wage for the time period in question will account for the effect of inflation. All material and supply costs have been adjusted to December 1975 using the Marshall and Stevens Index.

### 6.3.1 Sludge pumping

The cost of sludge pumping facilities presented in Figure 6-1 includes a positive displacement pump complete with motor and controls. No operation and maintenance costs are given because a sludge pump is an integral part of the dewatering system, and as a result the operation and maintenance costs of the system will cover the pump as well.

### 6.3.2 Gravity thickening

Equipment costs for gravity thickeners are presented in Figure 6-2, (Woods, 1976). These costs include a carbon steel tank complete with motor and drive, central rake, weirs and baffles. Painting of all components is also included. Operation and maintenance costs are assumed to be equal to those for a similar sized sedimentation basin and are presented in Figure 6-3.

### 6.3.3 Dissolved air flotation

The equipment costs presented in Figure 6-4 were obtained from manufacturers' quoted prices. The prices include the flotation basin, all required pumps and drives except the sludge feed pump, motors, instrumentation, air compressor, polymer feed system, and an electrical control panel. Operation and maintenance costs are shown in Figure 6-5.

### 6.3.4 Solid bowl scroll centrifuge

The equipment costs presented in Figure 6-6 were obtained from manufacturers' quoted prices. The costs include the centrifuge (stainless

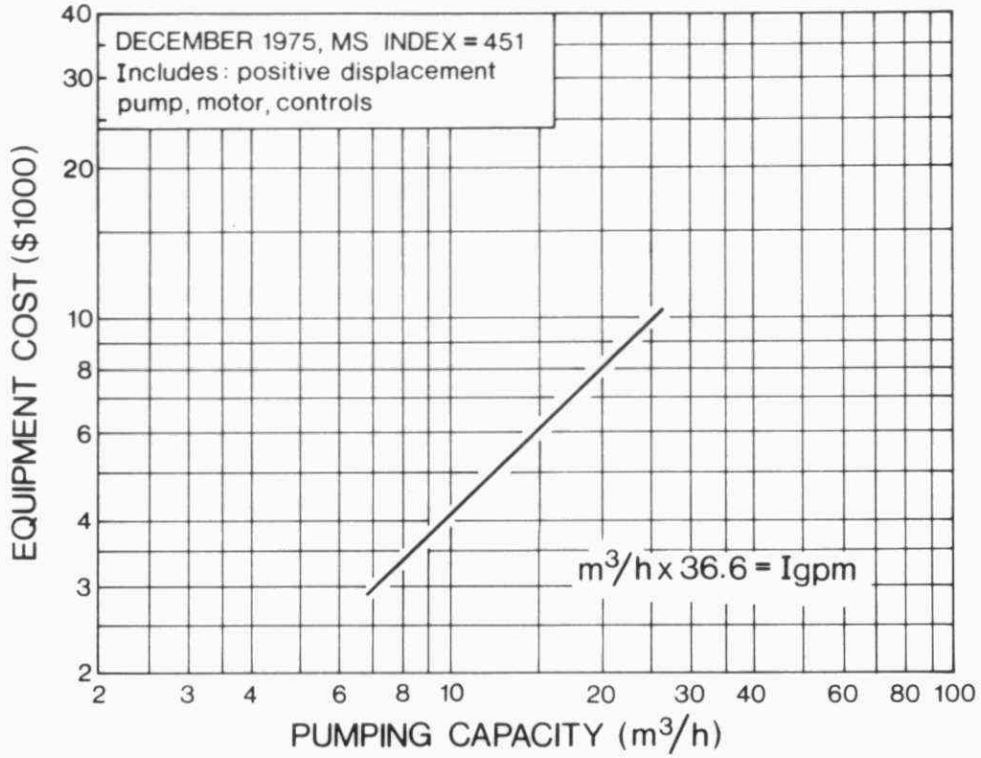


FIGURE 6-1. EQUIPMENT COST-SLUDGE PUMPS

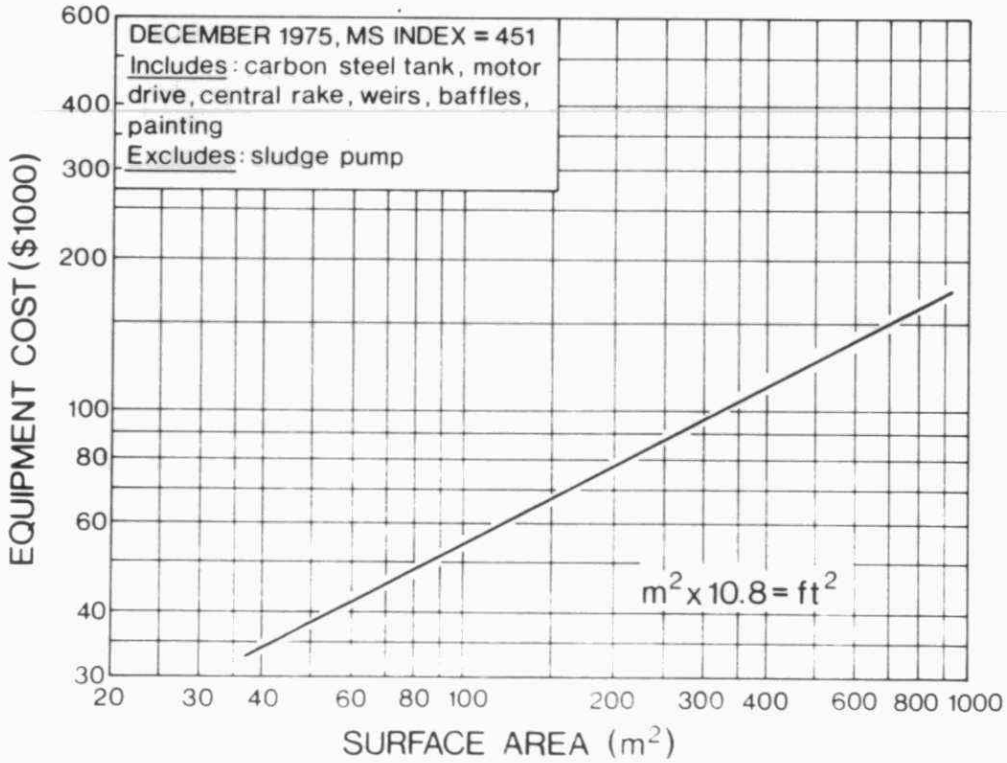


FIGURE 6-2. EQUIPMENT COST-GRAVITY THICKENERS

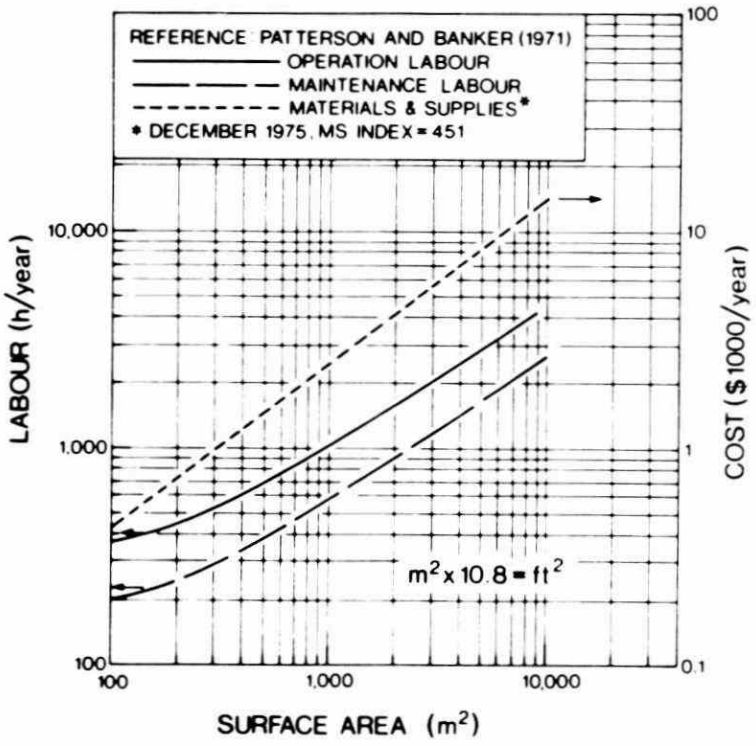


FIGURE 6-3. OPERATION AND MAINTENANCE COSTS-GRAVITY THICKENING

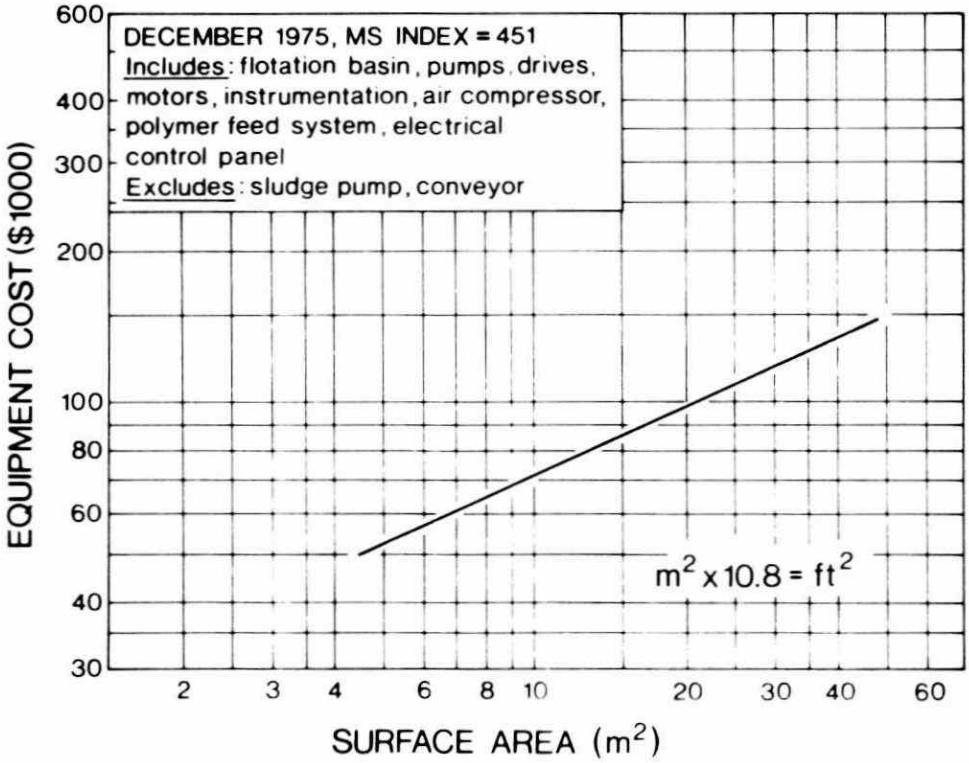


FIGURE 6-4. EQUIPMENT COST-DISSOLVED AIR FLOTATION

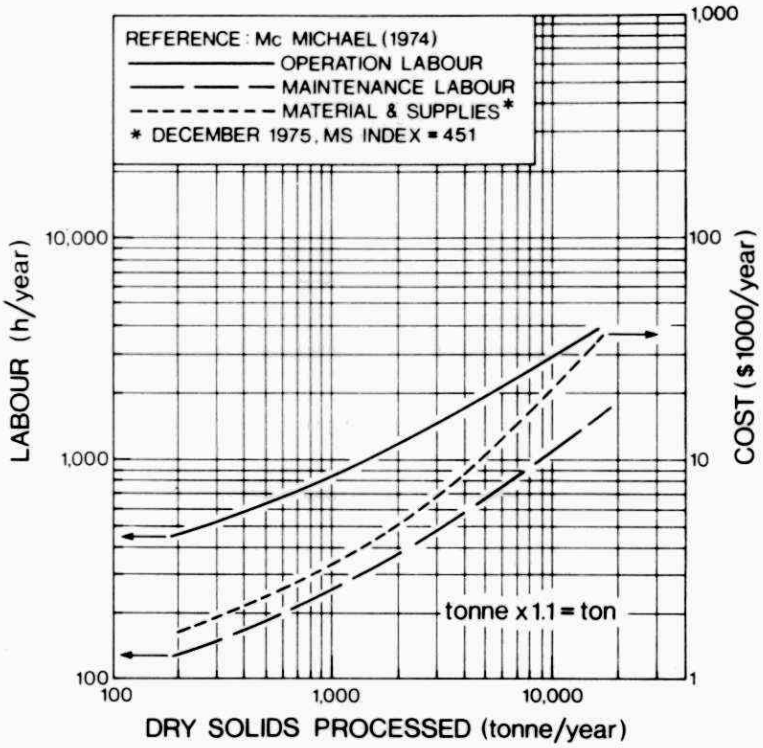


FIGURE 6-5. OPERATION AND MAINTENANCE COSTS - DISSOLVED AIR FLOTATION

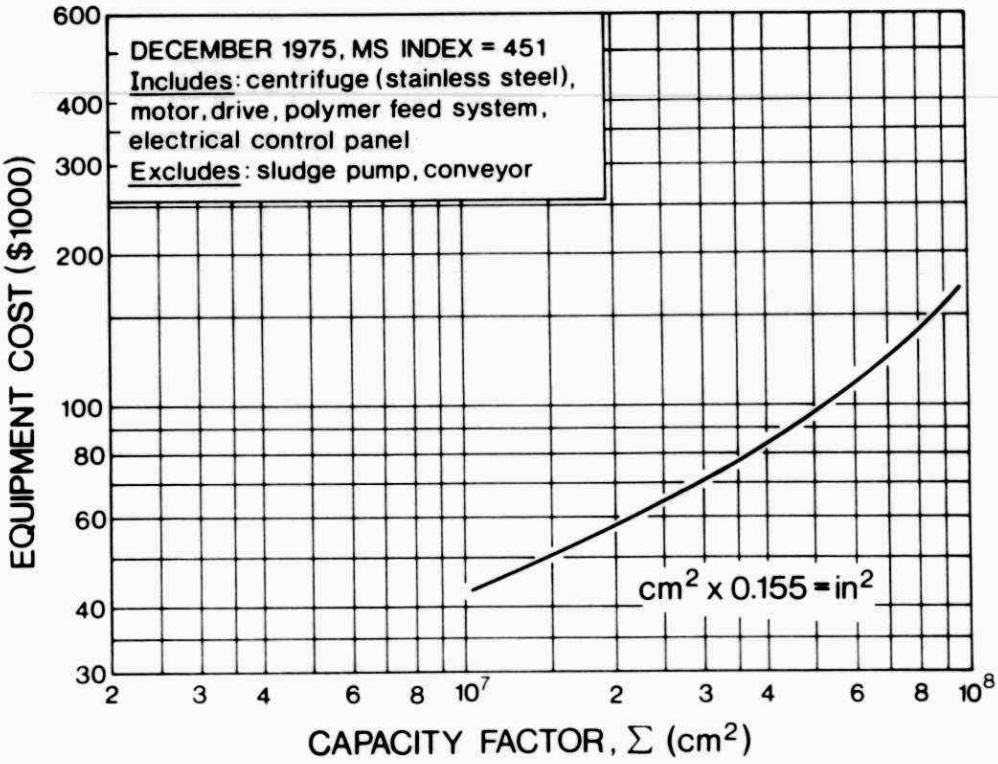


FIGURE 6-6. EQUIPMENT COST-SOLID BOWL CENTRIFUGE

steel construction), the motor, drive, polymer feed system and electrical control panel. Operation and maintenance costs are presented in Figure 6-7.

### 6.3.5 Imperforate bowl basket centrifuge

The equipment costs presented in Figure 6-8 were obtained from manufacturers' quoted prices. The costs include the centrifuge (stainless steel) motor and drive, polymer feed system, and electrical control panel. As shown in Figure 6-8, basket centrifuges suitable for wastewater sludge applications are only available in a limited range of sizes. No information has been presented on the operation and maintenance costs for basket centrifuges. It has been assumed that these costs can be approximated by the data presented in Figure 6-7 for solid bowl centrifuges.

### 6.3.6 Vacuum filtration

The equipment costs presented in Figure 6-9 were obtained from manufacturers' quoted prices. The costs include the rotary drum vacuum filter, fabric filter cloth, all pumps and drives excluding the sludge feed pump, motors, vacuum receiver, chemical feed system, and electrical control panel. Costs for a similar sized coil filter are 15% to 20% higher than for a fabric belt unit.

Operation and maintenance costs are presented in Figure 6-10.

### 6.3.7 Pressure filtration

The equipment costs presented in Figure 6-11 were obtained from manufacturers' quoted prices. The costs include the filter press, and a chemical feed system. Prices quoted are for cast iron construction recessed plate design, 2.54 cm (1") chambers and all equipment necessary for completely automatic operation of the filter press. Capital costs for polypropylene construction are competitive with those quoted for cast iron. Considerable equipment cost reduction can be obtained by selecting either semi-automatic (30% less) or manual operation (60% less). Operation and maintenance costs for automatic systems are presented in Figure 6-12.

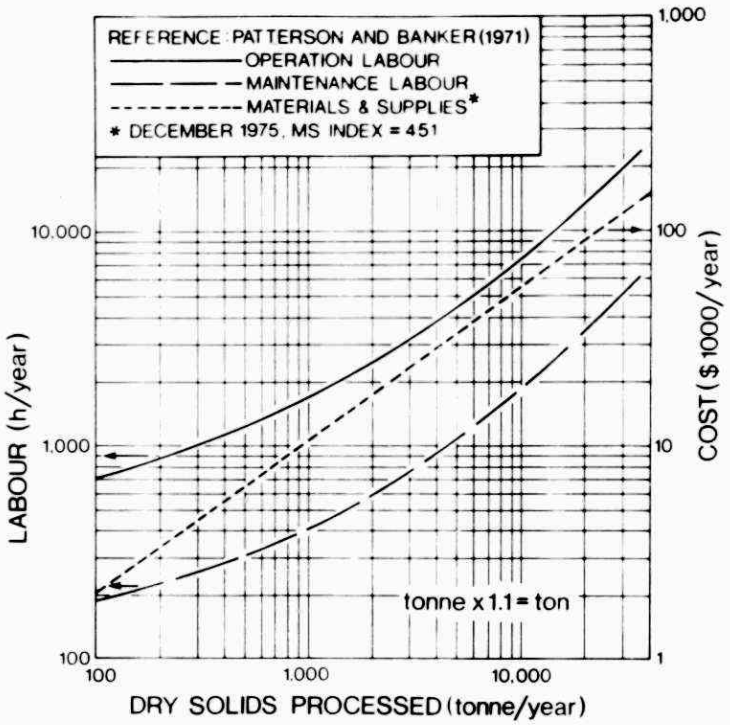


FIGURE 6-7. OPERATION AND MAINTENANCE COSTS-SOLID BOWL CENTRIFUGATION

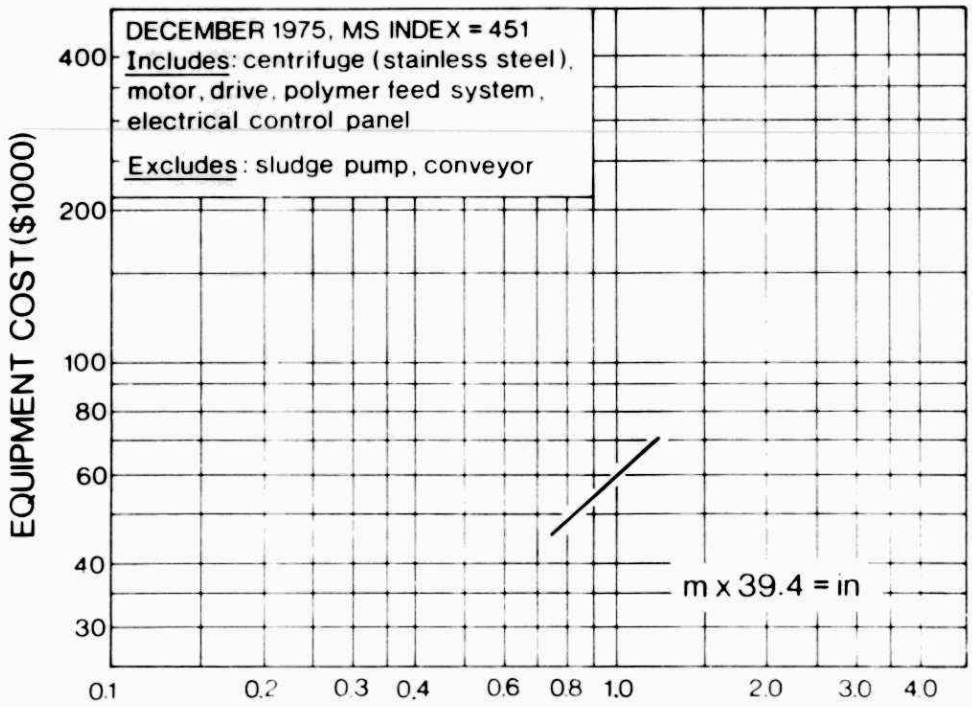


FIGURE 6-8. EQUIPMENT COST - BASKET CENTRIFUGE

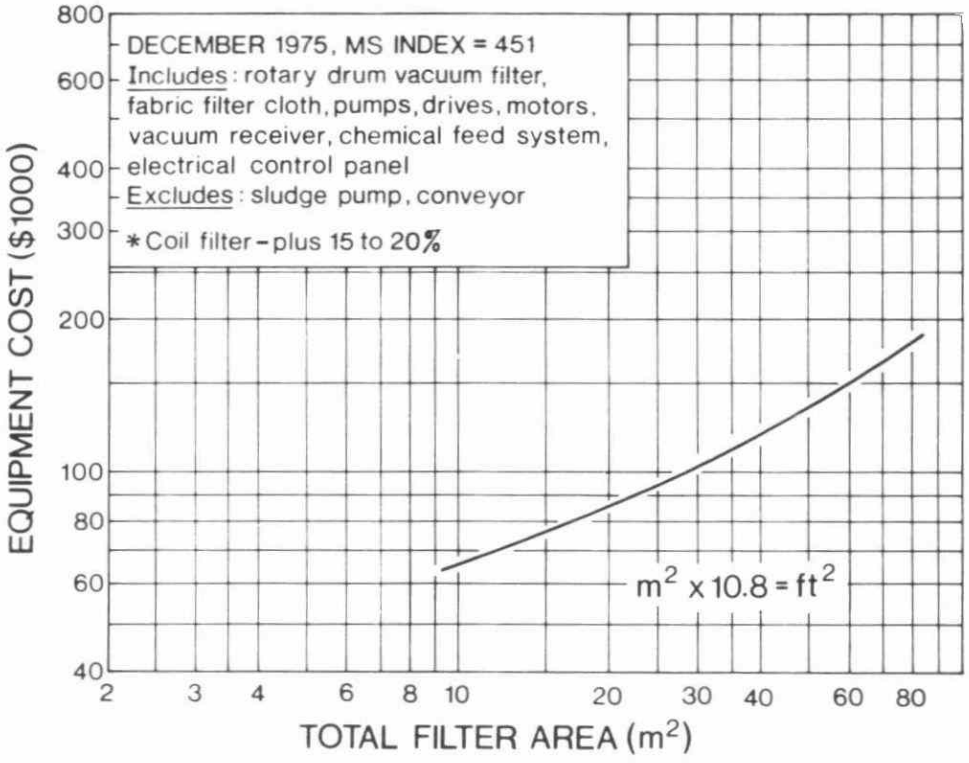


FIGURE 6-9. EQUIPMENT COST - VACUUM FILTER

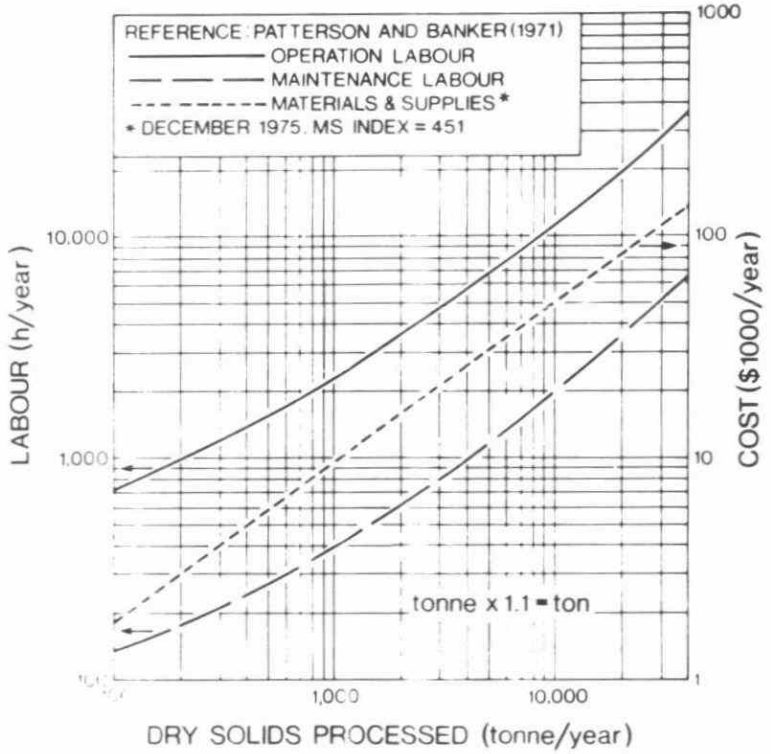


FIGURE 6-10. OPERATION AND MAINTENANCE COST-VACUUM FILTRATION

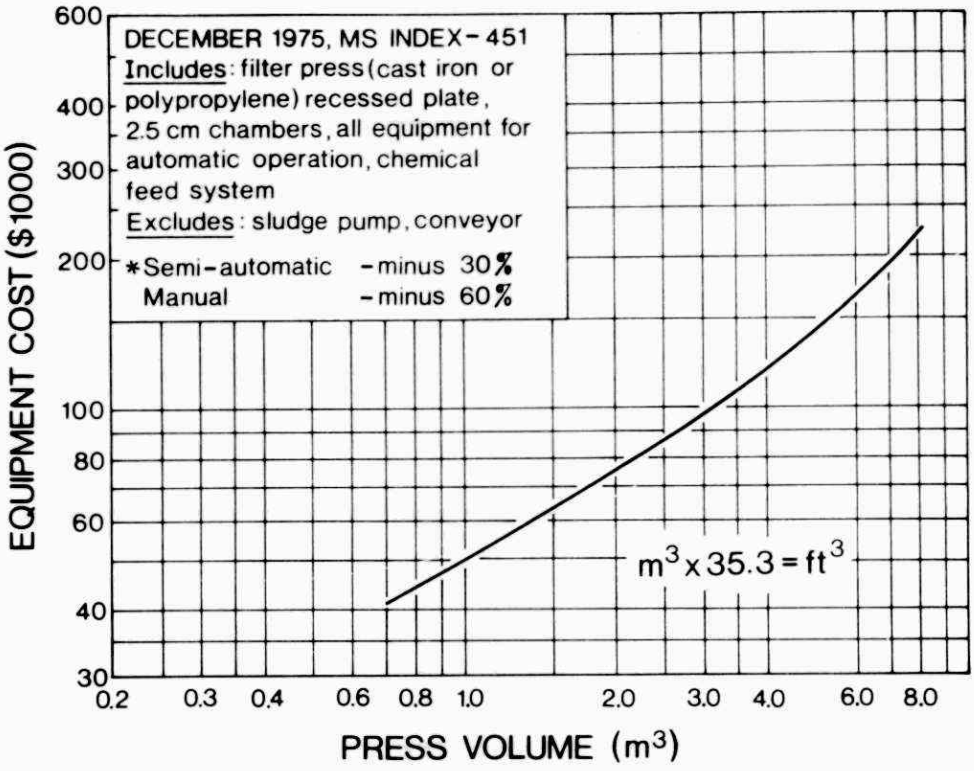


FIGURE 6-11. EQUIPMENT COST-PRESSURE FILTER

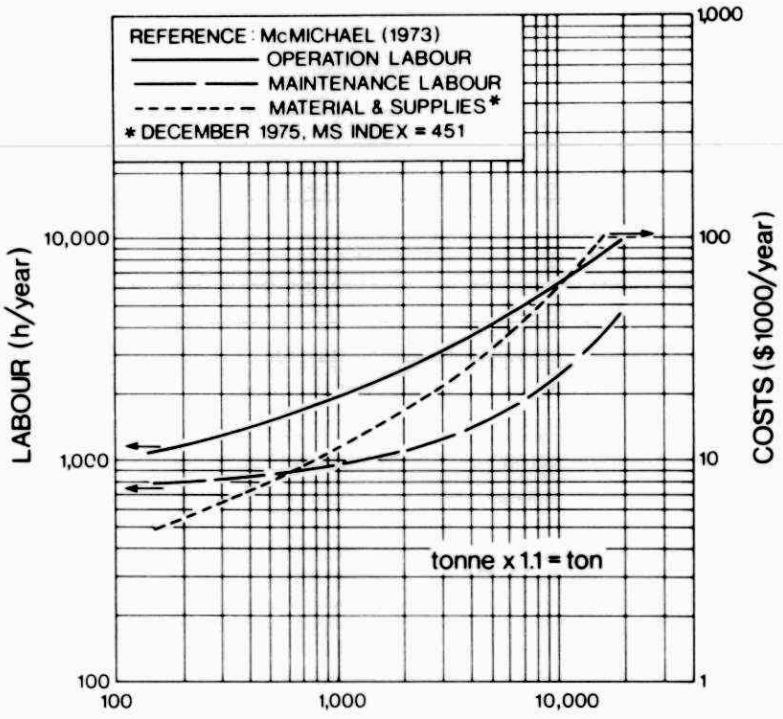


FIGURE 6-12. OPERATION AND MAINTENANCE COST-PRESSURE FILTRATION

### 6.3.8 Belt filter press

The equipment costs presented in Figure 6-13 were obtained from manufacturers' quoted prices. The costs include the belt press, one set of belts, all motors and drives, polymer feed system and electrical control panel. The width of that portion of the belt which is in contact with the sludge has been used as the design parameter. Generalized curves for operation and maintenance costs are not available. Estimates for belt life range from one to two years, and depend upon the feed material. The replacement cost for one complete set of belts varies from $1000 to $2000.

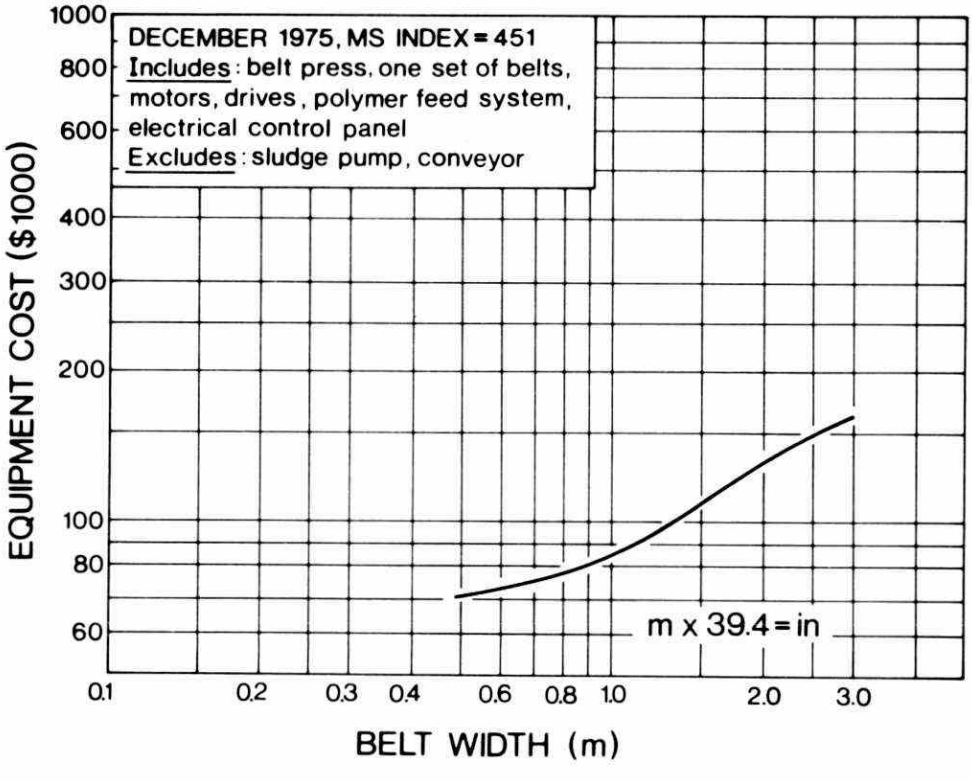


FIGURE 6-13. EQUIPMENT COST- BELT PRESS

## 6.4 Cost Estimate Examples

With the exception of the belt press, cost estimates have been developed for each dewatering process discussed in Section 5. In each case the estimate has been based on the information developed in the design example. Costs have been estimated on the basis of total cost per year and on cost per dry tonne of sludge solids. The cost estimate examples for gravity thickening, dissolved air flotation, solid bowl centrifugation, basket centrifugation, vacuum filtration and pressure filtration are summarized in Sections 6.4.1 through 6.4.6, respectively. The indirect costs which include engineering (10%), field expense (5%), legal and other (2%), contingency (6%), interest during construction (6%), and working capital (1.5%) have been shown in the examples as a single component equal to 30.5% of the contract cost.

6.4.1 Cost estimate example - gravity thickening

SLUDGE - design example - Section 5.1.3
- type - Waste Activated
- feed solids - 0.55%
- flow rate - 25 $m^3/h$ (90 Igpm)

DESIGN CAPACITY - sludge pump - 25 $m^3/h$ (90 Igpm)
- dewatering - surface area = 166 $m^2$ (1786 $ft^2$)

| | | | | |
|---|---|---|---|---|
| EQUIPMENT COST - sludge pump (Fig. 6-1) | $10,000 | | | |
| - dewatering (Fig. 6-2) | 70,000 | | | |
| | 80,000 | (1) | | |
| INSTALLATION FACTOR (TABLE 6-2) - 1.6 | | | | |
| INSTALLED COST | 1.6 x (1) | (2) | 128,000 | |
| TAX, FREIGHT AND INSURANCE | 8% of (1) | (3) | 6,400 | |
| TOTAL PHYSICAL PLANT COST | | (4) | 134,400 | |
| CONTRACTOR'S FEES | 12% of (4) | (5) | 16,130 | |
| CONTRACT COST | | (6) | 150,530 | |
| INDIRECT FEES | 30.5% of (6) | (7) | 45,910 | |
| CAPITAL COST | $ | (8) | 196,440 | |
| ANNUAL CAPITAL COST | 10.57% of (8) | $/year | | 20,760 |
| SLUDGE QUANTITY (Section 5.1.3) | | tonne/year | 1204 | |
| OPERATING LABOUR - FIGURE 6-3 | | h/year | 400 | |
| MAINTENANCE LABOUR - FIGURE 6-3 | | h/year | 220 | |
| UNIT LABOUR COST | | $/h | 6.30 | |
| ANNUAL LABOUR COST | | $/year | | 3,910 |
| ANNUAL MATERIAL AND SUPPLIES-FIGURE 6-3 | | $/year | | 590 |

| | Polymer | Lime | $FeCl_3$ |
|---|---|---|---|
| CHEMICAL DOSAGE (kg/tonne) | | | |
| CHEMICAL COST ($/kg) | | | |

| | | | | |
|---|---|---|---|---|
| TOTAL CHEMICAL COST | | $/tonne | | |
| ANNUAL CHEMICAL COST | | $/year | | - |
| TOTAL ANNUAL COST | | $/year | | 25,260 |
| TOTAL UNIT COST | | $/tonne | | 20.98 |

6.4.2 Cost estimate example - dissolved air flotation

SLUDGE - design example - Section 5.2.3
- type - Waste Activated
- feed solids - 1.0%
- flow rate - 25 $m^3/h$ (90 Igpm)

DESIGN CAPACITY - sludge pump - 25 $m^3/h$ (90 Igpm)
- dewatering - surface area = 14 $m^2$ (150 $ft^2$)

| | | | | |
|---|---|---|---|---|
| EQUIPMENT COST - sludge pump (Fig. 6-1) | $10,000 | | | |
| - dewatering (Fig. 6-4) | 84,000 | | | |
| | 94,000 | (1) | | |
| INSTALLATION FACTOR (TABLE 6-2) - 1.5 | | | | |
| INSTALLED COST | 1.5 x (1) | (2) | 141,000 | |
| TAX, FREIGHT AND INSURANCE | 8% of (1) | (3) | 7,520 | |
| TOTAL PHYSICAL PLANT COST | | (4) | 148,520 | |
| CONTRACTOR'S FEES | 12% of (4) | (5) | 17,820 | |
| CONTRACT COST | | (6) | 166,340 | |
| INDIRECT FEES | 30.5% of (6) | (7) | 50,730 | |
| CAPITAL COST | $ | (8) | 217,070 | |
| ANNUAL CAPITAL COST | 10.57% of (8) | $/year | | 22,940 |
| SLUDGE QUANTITY (Section 5.2.3) | | tonne/year | 2190 | |
| OPERATING LABOUR - FIGURE 6-5 | | h/year | 1200 | |
| MAINTENANCE LABOUR - FIGURE 6-5 | | h/year | 390 | |
| UNIT LABOUR COST | | $/h | 6.30 | |
| ANNUAL LABOUR COST | | $/year | | 10,020 |
| ANNUAL MATERIAL AND SUPPLIES-FIGURE 6-5 | | $/year | | 5,400 |

| | Polymer | Lime | $FeCl_3$ | | | |
|---|---|---|---|---|---|---|
| CHEMICAL DOSAGE (kg/tonne) | 1.85 | | | | | |
| CHEMICAL COST ($/kg) | 1.03 | | | | | |
| TOTAL CHEMICAL COST | | | | $/tonne | 1.91 | |
| ANNUAL CHEMICAL COST | | | | $/year | | 4,180 |
| TOTAL ANNUAL COST | | | | $/year | | 42,540 |
| TOTAL UNIT COST | | | | $/tonne | | 19.42 |

## 6.4.3 Cost estimate example - solid bowl scroll centrifuge

SLUDGE - design example - Section 5.3.3(a)
- type - Primary
- feed solids - 6%
- flow rate - 25 $m^3/h$ (90 lgpm)

DESIGN CAPACITY - sludge pump - 25 $m^3/h$ (90 lgpm)
- dewatering - capacity = $5.9 \times 10^7$ $cm^2$

| | | | | |
|---|---|---|---|---|
| EQUIPMENT COST - sludge pump (Fig. 6-1) | | $ 10,000 | | |
| - dewatering (Fig. 6-6) | | 110,000 | | |
| | | 120,000 | (1) | |
| INSTALLATION FACTOR (TABLE 6-2) - 1.6 | | | | |
| INSTALLED COST | | 1.6 x (1) | (2) | 192,000 |
| TAX, FREIGHT AND INSURANCE | | 8% of (1) | (3) | 9,600 |
| TOTAL PHYSICAL PLANT COST | | | (4) | 201,600 |
| CONTRACTOR'S FEES | | 12% of (4) | (5) | 24,190 |
| CONTRACT COST | | | (6) | 225,790 |
| INDIRECT FEES | | 30.5% of (6) | (7) | 68,870 |
| CAPITAL COST | | $ | (8) | 294,660 |
| ANNUAL CAPITAL COST | 10.57% of (8) | $/year | | 31,150 |
| SLUDGE QUANTITY (Section 5.3.3a) | | tonne/year | | 13,140 |
| OPERATING LABOUR - FIGURE 6-7 | | h/year | | 9,600 |
| MAINTENANCE LABOUR - FIGURE 6-7 | | h/year | | 2,400 |
| UNIT LABOUR COST | | $/h | | 6.30 |
| ANNUAL LABOUR COST | | $/year | | 75,600 |
| ANNUAL MATERIAL AND SUPPLIES-FIGURE 6-7 | | $/year | | 70,000 |

| | Polymer | Lime | $FeCl_3$ |
|---|---|---|---|
| CHEMICAL DOSAGE (kg/tonne) | 2.5 | | |
| CHEMICAL COST ($/kg) | 1.03 | | |

| | | | |
|---|---|---|---|
| TOTAL CHEMICAL COST | $/tonne | 2.58 | |
| ANNUAL CHEMICAL COST | $/year | | 33,900 |
| TOTAL ANNUAL COST | $/year | | 210,650 |
| TOTAL UNIT COST | $/tonne | | 16.03 |

6.4.4 Cost estimate example - imperforate bowl basket centrifuge

SLUDGE - design example - Section 5.3.4(b)
- type - Waste Activated
- feed solids - 0.9%
- flow rate - 25 $m^3/h$ (90 Igpm)

DESIGN CAPACITY - sludge pump - 2 at 12.5 $m^3/h$ (45 Igpm)
- dewatering - 2 at 1.2 m diameter (48 in)

| | | | | | |
|---|---|---|---|---|---|
| EQUIPMENT COST - sludge pump (Fig. 6-1) | | $ 10,000 | | | |
| - dewatering (Fig. 6-8) | | 138,000 | | | |
| | | 148,000 | (1) | | |
| INSTALLATION FACTOR (TABLE 6-2) - 1.57 | | | | | |
| INSTALLED COST | | 1.57 x (1) | (2) | 232,360 | |
| TAX, FREIGHT AND INSURANCE | | 8% of (1) | (3) | 11,840 | |
| TOTAL PHYSICAL PLANT COST | | | (4) | 244,200 | |
| CONTRACTOR'S FEES | | 12% of (4) | (5) | 29,300 | |
| CONTRACT COST | | | (6) | 237,500 | |
| INDIRECT FEES | | 30.5% of (6) | (7) | 83,420 | |
| CAPITAL COST | | $ | (8) | 320,920 | |
| ANNUAL CAPITAL COST | 10.57% of (8) | $/year | | | 33,920 |
| SLUDGE QUANTITY (Section 5.3.4b) | | tonne/year | | 1971 | |
| OPERATING LABOUR - FIGURE 6-7 | | h/year | | 2500 | |
| MAINTENANCE LABOUR - FIGURE 6-7 | | h/year | | 580 | |
| UNIT LABOUR COST | | $/h | | 6.30 | |
| ANNUAL LABOUR COST | | $/year | | | 19,400 |
| ANNUAL MATERIAL AND SUPPLIES - FIGURE 6-7 | | $/year | | | 17,000 |

| | Polymer | Lime | $FeCl_3$ |
|---|---|---|---|
| CHEMICAL DOSAGE (kg/tonne) | 1.25 | | |
| CHEMICAL COST ($/kg) | 1.03 | | |

| | | | | | |
|---|---|---|---|---|---|
| TOTAL CHEMICAL COST | | $/tonne | | 1.29 | |
| ANNUAL CHEMICAL COST | | $/year | | | 2,540 |
| TOTAL ANNUAL COST | | $/year | | | 72,860 |
| TOTAL UNIT COST | | $/tonne | | | 36.97 |

## 6.4.5 Cost estimate example - vacuum filtration

SLUDGE - design example - Section 5.4.3
- type - Digested
- feed solids - 5.7%
- flow rate - 25 $m^3/h$ (90 Igpm)

DESIGN CAPACITY - sludge pump - 25 $m^3/h$ (90 Igpm)
- dewatering - surface area = 59.4 $m^2$ (640 $ft^2$)

| | | | | |
|---|---|---|---|---|
| EQUIPMENT COST - sludge pump (Fig. 6-1) | | $ 10,000 | | |
| - dewatering (Fig. 6-9) | | 145,000 | | |
| | | 155,000 | (1) | |
| INSTALLATION FACTOR (TABLE 6-2) - 1.8 | | | | |
| INSTALLED COST | | 1.8 x (1) | (2) | 279,000 |
| TAX, FREIGHT AND INSURANCE | | 8% of (1) | (3) | 12,400 |
| TOTAL PHYSICAL PLANT COST | | | (4) | 291,400 |
| CONTRACTOR'S FEES | | 12% of (4) | (5) | 34,970 |
| CONTRACT COST | | | (6) | 326,370 |
| INDIRECT FEES | | 30.5% of (6) | (7) | 99,540 |
| CAPITAL COST | | $ | (8) | 425,910 |
| ANNUAL CAPITAL COST | 10.57% of (8) | $/year | | 45,020 |
| SLUDGE QUANTITY (Section 5.4.3) | | tonne/year | | 12,483 |
| OPERATING LABOUR - FIGURE 6-10 | | h/year | | 13,000 |
| MAINTENANCE LABOUR - FIGURE 6-10 | | h/year | | 2,300 |
| UNIT LABOUR COST | | $/h | | 6.30 |
| ANNUAL LABOUR COST | | $/year | | 96,390 |
| ANNUAL MATERIAL AND SUPPLIES-FIGURE 6-10 | | $/year | | 60,000 |

| | Polymer | Lime | $FeCl_3$ | | | |
|---|---|---|---|---|---|---|
| CHEMICAL DOSAGE (kg/tonne) | | 90 | 60 | | | |
| CHEMICAL COST ($/kg) | | 0.044 | 0.190 | | | |
| TOTAL CHEMICAL COST | | | | $/tonne | | 15.36 |
| ANNUAL CHEMICAL COST | | | | $/year | | 191,740 |
| TOTAL ANNUAL COST | | | | $/year | | 393,150 |
| TOTAL UNIT COST | | | | $/tonne | | 31.49 |

## 6.4.6 Cost estimate example - pressure filtration

SLUDGE - design example - Section 5.5.3
- type - Acid Mine Drainage
- feed solids - 7.0%
- flow rate - 25 $m^3/h$ (90 Igpm)

DESIGN CAPACITY - sludge pump - 25 $m^3/h$ (90 Igpm)
- dewatering - press volume = 2.7 $m^3$ (95 $ft^3$)

| | | | | | |
|---|---|---|---|---|---|
| EQUIPMENT COST - sludge pump (Fig. 6-1) | | $10,000 | | | |
| - dewatering (Fig. 6-11) | | 90,000 | | | |
| | | 100,000 | (1) | | |
| INSTALLATION FACTOR (TABLE 6-2) - 1.79 | | | | | |
| INSTALLED COST | | 1.79 x (1) | (2) | 179,000 | |
| TAX, FREIGHT AND INSURANCE | | 8% of (1) | (3) | 8,000 | |
| TOTAL PHYSICAL PLANT COST | | | (4) | 187,000 | |
| CONTRACTOR'S FEES | | 12% of (4) | (5) | 22,440 | |
| CONTRACT COST | | | (6) | 209,440 | |
| INDIRECT FEES | | 30.5% of (6) | (7) | 63,880 | |
| CAPITAL COST | | $ | (8) | 273,320 | |
| ANNUAL CAPITAL COST | 10.57% of (8) | $/year | | | 28,890 |
| SLUDGE QUANTITY (Section 5.5.3) | | tonne/year | | 15,987 | |
| OPERATING LABOUR - FIGURE 6-12 | | h/year | | 8,500 | |
| MAINTENANCE LABOUR - FIGURE 6-12 | | h/year | | 3,700 | |
| UNIT LABOUR COST | | $/h | | 6.30 | |
| ANNUAL LABOUR COST | | $/year | | | 76,860 |
| ANNUAL MATERIAL AND SUPPLIES-FIGURE 6-12 | | $/year | | | 100,000 |

| | Polymer | Lime | $FeCl_3$ |
|---|---|---|---|
| CHEMICAL DOSAGE (kg/tonne) | | | |
| CHEMICAL COST ($/kg) | | | |

| | | | | | |
|---|---|---|---|---|---|
| TOTAL CHEMICAL COST | | $/tonne | | | |
| ANNUAL CHEMICAL COST | | $/year | | | - |
| TOTAL ANNUAL COST | | $/year | | | 205,750 |
| TOTAL UNIT COST | | $/tonne | | | 12.87 |

The techniques outlined in this manual should not be considered to represent the only acceptable, or necessarily the best methods of designing a particular piece of sludge dewatering equipment. In some instances, especially with specialized equipment, manufacturers may have developed more comprehensive design procedures which they do not publish due to their proprietary nature. The limitations of preparing cost estimates based on generalized cost curves has also been emphasized. The design techniques and cost data presented do allow a designer to arrive at a reasonable estimate of his dewatering requirements, and to prepare preliminary cost estimates for comparing those processes which are technically feasible.

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# APPENDIX

## COST FORMULAE

All of the cost curves in Chapter 6 can be represented by one of the following models:

$$\log y = a \log x + b \qquad \text{(A-1)}$$

or

$$\log y = \frac{1}{a \log x + b} \qquad \text{(A-2)}$$

The values for x, y, a, and b, and the boundary limits (in terms of x) for each curve are presented in Table A-1. The units for x and y must be as shown in the table, in order for the values of the constants (a and b) to be valid. It must be stressed that any attempt to use either equation (A-1 or A-2) outside the boundary conditions listed, may result in extremely large errors.

TABLE A-1. COST FORMULAE

| Figure | Unit Process | Equation | y | x | a | b | Boundary Limits |
|---|---|---|---|---|---|---|---|
| 6-1 | Sludge Pumps | A-1 | Equipment Cost ($1,000) | Pumping Capacity ($m^3/h$) | 0.953 | -0.345 | 7 - 26 $m^3/h$ |
| 6-2 | Gravity Thickeners | A-1 | Equipment Cost ($1,000) | Surface Area ($m^2$) | 0.507 | 0.725 | 37 - 900 $m^2$ |
| 6-3 | Gravity Thickening (Operation Labour) | A-2 | Labour (h/year) | Surface Area ($m^2$) | -0.056 | 0.502 | 100 - 5,000 $m^2$ |
| 6-3 | Gravity Thickening (Maintenance Labour) | A-2 | Labour (h/year) | Surface Area ($m^2$) | -0.075 | 0.585 | 100 - 5,000 $m^2$ |
| 6-3 | Gravity Thickening (Materials and Supplies) | A-1 | Cost ($1,000/year) | Surface Area ($m^2$) | 0.756 | -1.877 | 100 - 10,000 $m^2$ |
| 6-4 | Dissolved Air Flotation | A-1 | Equipment Cost ($1,000) | Surface Area ($m^2$) | 0.448 | 1.409 | 4.4 - 47 $m^2$ |
| 6-5 | Dissolved Air Flotation (Operation Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.051 | 0.493 | 200 - 15,000 tonne/year |
| 6-5 | Dissolved Air Flotation (Maintenance Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.084 | 0.666 | 200 - 15,000 tonne/year |
| 6-5 | Dissolved Air Flotation (Materials and Supplies) | A-2 | Cost ($1,000/year) | Solids Processed (tonne/year) | -1.359 | 6.006 | 500 - 6,000 tonne/year |
| 6-6 | Solid Bowl Centrifuge | A-2 | Equipment Cost ($1,000) | Capacity, $\Sigma$ ($cm^2$) | -0.168 | 1.790 | $10^7$ - $10^8$ $cm^2$ |
| 6-7 | Solid Bowl Centrifuge (Operation Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.047 | 0.446 | 100 - 30,000 tonne/year |

TABLE A-1. COST FORMULAE (CONT'D)

| Figure | Unit Process | Equation | y | x | a | b | Boundary Limits |
|---|---|---|---|---|---|---|---|
| 6-7 | Solid Bowl Centrifuge (Maintenance Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.067 | 0.574 | 100 - 30,000 tonne/year |
| 6-7 | Solid Bowl Centrifuge (Material and Supplies) | A-1 | Cost ($1,000/year) | Solids Processed (tonne/year) | 0.724 | -1.146 | 100 - 30,000 tonne/year |
| 6-8 | Basket Centrifuge | A-1 | Equipment Cost ($1,000) | Basket Diameter (m) | 0.405 | 1.353 | 0.76 - 1.2 m |
| 6-9 | Vacuum Filter | A-2 | Equipment Cost ($1,000) | Total Filter Area ($m^2$) | -0.122 | 0.674 | 9.5 - 84 $m^2$ |
| 6-10 | Vacuum Filtration (Operation Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.052 | 0.455 | 100 - 40,000 tonne/year |
| 6-10 | Vacuum Filtration (Maintenance Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.083 | 0.635 | 100 - 40,000 tonne/year |
| 6-10 | Vacuum Filtration (Material and Supplies) | A-1 | Cost ($1,000/year) | Solids Processed (tonne/year) | 0.722 | -1.189 | 100 - 40,000 tonne/year |
| 6-11 | Pressure Filter | A-2 | Equipment Cost ($1,000) | Press Volume ($m^3$) | -0.179 | 0.588 | 0.7 - 8.0 $m^3$ |
| 6-12 | Pressure Filtration (Operation Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.036 | 0.408 | 150 - 15,000 tonne/year |
| 6-12 | Pressure Filtration (Maintenance Labour) | A-2 | Labour (h/year) | Solids Processed (tonne/year) | -0.011 | 0.369 | 150 - 6,000 tonne/year |
| 6-12 | Pressure Filtration (Material and Supplies) | A-2 | Cost ($1,000/year) | Solids Processed (tonne/year) | -0.399 | 2.158 | 400 - 15,000 tonne/year |
| 6-13 | Belt Press | A-2 | Equipment Cost ($1,000) | Belt Width (m) | -0.103 | 0.507 | 0.5 - 3.0 m |

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